Opportunities for solar process heat integration and heat recovery in the South African fishmeal industry

by

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Abstract

Solar thermal renewable energy is a promising alternative heat source capable of providing a large portion of the South African industrial heat demand. The major energy demand within the energy intensive South African industrial sector is process heat, furthermore, industrial process heat constitutes approximately 30% of the national annual energy consumption. Most of this heat is currently supplied by fossil fuels, which is a challenge to the future sustainability of the industrial sector since the cost of fossil fuels is expected to increase indefinitely, and their use impacts negatively on the environment.

Two South African fishmeal factories were studied with the aim of determining the feasibility of integrating solar thermal heat into existing production processes within the industrial sector. The fishmeal production process is energy intensive as it requires the evaporation of large amounts of water.

Base case processes were established, based on actual production data collected from the factories, in order to determine the energy and fuel requirements of the factories. Opportunities for heat recovery and solar heat integration were identified, and their effects on the energy demand quantified. The total potential for solar heat (in terms of total collector area) was established and two systems proposed: 1) with an area that minimised the difference between solar heat demand and supply, and 2) with an area that resulted in no excess heat production. A preliminary economic analysis was performed to quantify the economic viability of the proposed systems.

Factory A produces fishmeal from lean-fish processing by-products using a single dryer, with heavy fuel oil as fuel source. Preheating of the raw material stream presented an opportunity for both solar heat integration and heat recovery. A 384 m^2 solar heat system was the most profitable option investigated with a net present value of R 3.3 million and levelized cost of heat of R 0.79. Heat recovery from the condensate stream exiting the dryer was also economically viable, however, it was less profitable and resulted in lower fuel savings.

Factory B produces fishmeal and fish oil from pelagic fish species using the wet-pressing method, with coal as fuel source. Solar thermal heat could be used to preheat the entering raw material and boiler make-up water streams and to heat the stickwater concentrate prior to drying. Heat recovery from the fish oil stream could only supply a very small fraction of the heat required. Due to the large capital costs of the solar thermal systems and the low cost of coal, none of the proposed systems were economically viable.

The cost of the fuel being replaced and the heat demand throughout the year were found to be major factors affecting the economic viability of the solar thermal heat systems. It is recommended that the energy requirements and production schedules determined in this study, be used to simulate the solar heat systems and obtain more accurate values of the solar thermal system efficiency and output. This will aid the specific factories to obtain implementable solutions.

Opsomming

Hernubare sonverhittings energie is 'n belowende alternatiewe hitte bron wat 'n groot gedeelte van die Suid-Afrikaanse industriële hitte vraag kan voorsien. Die grootste vraag vir energie in die energie intensiewe Suid-Afrikaanse industriële sektor is vir proses hitte, verder maak industriële proses hitte ongeveer 30% van die nasionale jaarlikse energie verbruik uit. Die meeste van die hitte word tans deur fossiel brandstowwe voorsien, wat 'n uitdaging is vir die toekomstige volhoubaarheid van die industriële sektor, siende dat die koste van fossiel brandstowwe verwag word om onbepaald toe te neem, en die gebruik daarvan 'n negatiewe impak op die omgewing het.

Twee Suid-Afrikaanse vismeel fabrieke was bestudeer met die doel om die lewensvatbaarheid van die insluiting van sonverhitting in bestaande produksie prosesse binne die industriële sektor te bepaal. Die vismeel produksie proses is energie intensief weens die feit dat dit die verdamping van groot hoeveelhede water vereis.

Basis geval prosesse was gestig, gebaseer op werklike produksie data wat by die fabrieke ingesamel was, om die energie vereistes en brandstof verbruik van die fabrieke te bepaal. Geleenthede vir hitte herwinning en die insluiting van sonverhitting was geïdentifiseer en die effekte daarvan op die energie vraag gekwantifiseer. Die totale potensiaal vir sonverhitting (in terme van die totale versamelaar area) was bepaal en twee sisteme voorgestel: 1) met 'n area wat die verskil tussen die hitte vraag en aanbod minimeer het, en 2) met 'n area wat geen ongebruikte hitte tot gevolg gehad het nie. 'n Voorlopige ekonomiese analise was uitgevoer om die ekonomiese lewensvatbaarheid van die voorgestelde sisteme te bepaal.

Fabriek A produseer vismeel vanaf maer-vis prosessering byprodukte met 'n enkele droër, met swaar olie as brandstof. Voorafverhitting van die rou materiaal stroom het 'n geleentheid gebied vir beide sonverhitting en hitte herwinning. 'n 384 m² sonverhittingstelsel was die mees winsgewende opsie wat ondersoek was, met 'n netto huidige waarde van R 3.3 miljoen en 'n genormaliseerde hitte koste van R 0.79. Hitte herwinning vanaf die kondensaat stroom wat die droër verlaat was ook ekonomies lewensvatbaar, dit was egter minder winsgewend en het minder brandstof besparings tot gevolg gehad.

Fabriek B produseer vismeel en vis olie vanaf pelagiese vis spesies met die nat-druk metode, met steenkool as brandstof. Sonverhitting kan gebruik word om die rou materiaal stroom, die addisionele ketel water, en die konsentraat te verhit. Hitte herwinning vanaf die vis olie stroom kon slegs 'n baie klein gedeelte van die vereiste hitte voorsien. Weens die groot kapitaal koste van die sonverhittingstelsels en die lae koste van steenkool, was geen van die voorgestelde stelsels vir Fabriek B ekonomies lewensvatbaar nie.

Die koste van die brandstof wat vervang word en die hitte vraag deur die loop van die jaar het die grootste effek op die ekonomiese lewensvatbaarheid van die sonverhittingstelsels gehad. Dit word aanbeveel dat die hitte vereistes en produksie skedules wat in die studie bepaal was, gebruik word om die sonverhittingstelsels te simuleer en sodoende meer akkurate waardes van die sisteem doeltreffendheid en uitset te kry. Dit sal die spesifieke fabrieke help om 'n implementeerbare oplossing te vind.

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Abbreviations

BFD	Block Flow Diagram
CPC	Compound Parabolic Concentrator
ETC	Evacuated Tube Collector
FA	Factory A
FAQ	Fair Average Quality
FB	Factory B
FMFO	Fishmeal and Fish Oil
FPC	Flat Plate Collector
IRR	Internal Rate of Return
LCOH	Levelized Cost of Heat
LFR	Linear Fresnel Reflector
NPV	Net Present Value
PFD	Process Flow Diagram
PLC	Programmable Logic Controller
PTC	Parabolic Trough Concentrator
SA-STTRM	South African Solar Thermal Technology Roadmap
SETRM	Solar Energy Technology Roadmap
SF	Solar Fraction
SHIP	Solar Heat for Industrial Processes
ST	Solar Thermal
STC	Solar Thermal Collector
TMY	Typical Meteorological Year

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1 <u>Introduction</u>

In this chapter a brief background is presented which substantiates the motivation for this study. The scope of this study is defined, followed by the aim and objectives. Lastly, the document structure and content are stated.

1.1 Background

The South African industrial sector is energy intensive, with 65% of the energy demand in this sector being destined for process heat (Hess, 2016b). Fossil fuels are the main sources of heat in the South African industrial sector (Joubert et al., 2016). The continued use of fossil fuels is problematic since their finite nature means their cost will increase indefinitely (Joubert et al., 2016), and their use is damaging the environment (Thirugnanasambandam et al., 2010).

The fishmeal and fish oil production industry in the Western Cape is important to South Africa. It supports the economy and provides employment to thousands in rural and economically underdeveloped regions, either directly in processing factories or through the pelagic fishery (Hara et al., 2008). Fishmeal production is energy intensive, as it requires the evaporation of large amounts of water (Windsor, 2001). Steam is typically used as energy carrier in fishmeal factories, generated in boilers that combust fossil fuels such as coal and heavy fuel oil.

Renewable energy sources aid in reducing the dependence on fossil fuels, and the negative environmental impact associated with their use (Tian and Zhao, 2013). Solar energy is the renewable resource with the greatest potential in South Africa (Pegels, 2010), due to a solar resource with high levels of irradiation over large parts of the country (WWF, 2017). Various technologies already exploit solar energy, of which domestic hot water production through solar thermal heating is an example. A further promising application of solar thermal technology is to produce heat for use in industrial processes.

Two South African fishmeal factories were studied to identify opportunities for integration of solar thermal heat into existing production operations. Both factories are situated along the western coast of South Africa, where most of the fishmeal industry is located (Hara et al., 2008). Production at the two factories is distinct from each other, varying in the type of raw material used, the production capacity and rate, and the production schedule throughout the year. Mass and energy balances based on real plant operating data were used to create simulations of the two factories in Microsoft Excel 2016 and Aspen Plus V8.8, characterising the operation of the plant during typical operating conditions. These simulations were used to quantify the effect of heat recovery or solar thermal heat

integration on the energy requirements and fuel consumption of the factories. The feasibility of the different integration options proposed was quantified with a preliminary economic analysis.

1.2 Study motivation and scope

1.2.1 Motivation for the study

Solar thermal heat is a promising technology that uses freely available solar radiation as an energy source. Solar thermal better utilises the available solar irradiation than other solar energy technologies, since solar thermal collectors typically operate at significantly higher efficiencies than solar photovoltaic collectors (Tian and Zhao, 2013). This study sought to identify opportunities for solar thermal heating within the South African industrial sector, as this has the potential to significantly reduce the energy required from fossil fuels to produce heat. The fishmeal production industry was selected as a case study since it is known to be an energy intensive industry, and there is no information readily available on the energy consumption of fishmeal and fish oil factories in South Africa; furthermore, the potential for solar process heat in this industry is also unknown.

This study contributes to the solar heat for industrial processes knowledge base, and the fishmeal and fish oil industry knowledge base. Information regarding the feasibility of integrating solar thermal heat into existing fishmeal factories located in areas of relatively high solar irradiation would be valuable. By using recent real plant data for different factories with high energy demand in this study the results are applicable to industry.

1.2.2 Scope of the study

The scope of this study encompassed two main aspects: describing production at South African fishmeal factories and using mass and energy balances based on real plant data to identify opportunities to reduce energy required from fossil fuels, followed by simulating the effects of different heat integration proposals on the energy requirements of the factory and determining the economic viability of the proposed solar heat and heat recovery systems. The study culminated in reporting the production process and the results of the solar heat integration study in this document.

1.3 Aim and objectives

The aim of this study was to determine the feasibility of integrating solar thermal heat into existing production processes, with fishmeal factories located in South Africa selected as case studies due to the high energy-demand of these processes and relatively high solar irradiation at their locations.

The following objectives were defined:

i. Define and characterise base case processes for two South African fishmeal factories using actual production and energy usage data collected from the factories.

ii. Compare the performance of the factories against international industry standards and determine if the production processes of South African factories deviate from international practice.

iii. Identify opportunities for conventional energy efficiency measures, and for solar heat integration in the factories, in order to quantify the effects of implementing these changes on the heat and fuel requirements of the studied production plants.

iv. Identify a suitable solar thermal collector technology to be used and determine the total solar thermal collector area required to supply the total heat demand. Also investigate other solar thermal collector areas to maximise the amount of utilisable heat obtained from the solar heat system.

v. Perform a preliminary economic analysis to quantify the economic viability of the proposed solar heat integration options.

1.4 Document structure

The document is structured around five chapters and deviates from the conventional structure where all the literature, methods and results are grouped into individual chapters. Describing the production process of South African fishmeal factories and investigating the feasibility of integrating solar thermal heat into existing production processes are two distinct endeavours. Therefore, the results of these distinct parts are presented in separate chapters, along with the more detailed literature and methods relevant to each aspect. To understand the context and implication of the results, a general overview of the South African energy demand, solar thermal heating and fishmeal manufacture is provided early in the document.

Chapter One provides a brief introduction to the study, stating the motivation, scope, and aim and objectives. Chapter Two provides a general introduction and literature overview of the relevant topics: the South African energy demand, solar thermal technologies and the most commonly used fishmeal production process. Chapters Three and Four contain the more detailed aspects of the fishmeal production process and solar thermal heat integration, respectively, along with the methods used, and results relevant to each aspect. Chapter Five provides the conclusions and recommendations arising from the study.

2   Context and literature overview

In this chapter an overview of the South African energy demand is given, the fuels currently used and their environmental impact are discussed, along with the solar resource available in South Africa. Aspects around solar thermal collectors, and the solar heat market, both internationally and in South Africa, are discussed. The properties and production of fishmeal and fish oil are discussed, and the importance of the fishmeal industry to the South African economy is highlighted.

2.1   South African energy landscape

2.1.1   Energy demand per economic sector

The industrial sector in South Africa is responsible for the largest proportion of national energy consumption, with most of the energy being used to produce process heat. The share of total energy demand per economic sector is shown in Figure 2-1 [Left]; the industrial sector consumes the most at 46%, followed by transportation with a share of 29%, the agricultural sector consumes the least at 3% of the total energy demand. Figure 2-1 [Right] shows the shares of energy demand within the South African industrial sector; the total process heat demand sums to 65% of industrial energy demand which amounts to almost 30% of the total national energy consumed annually (Hess, 2016b).
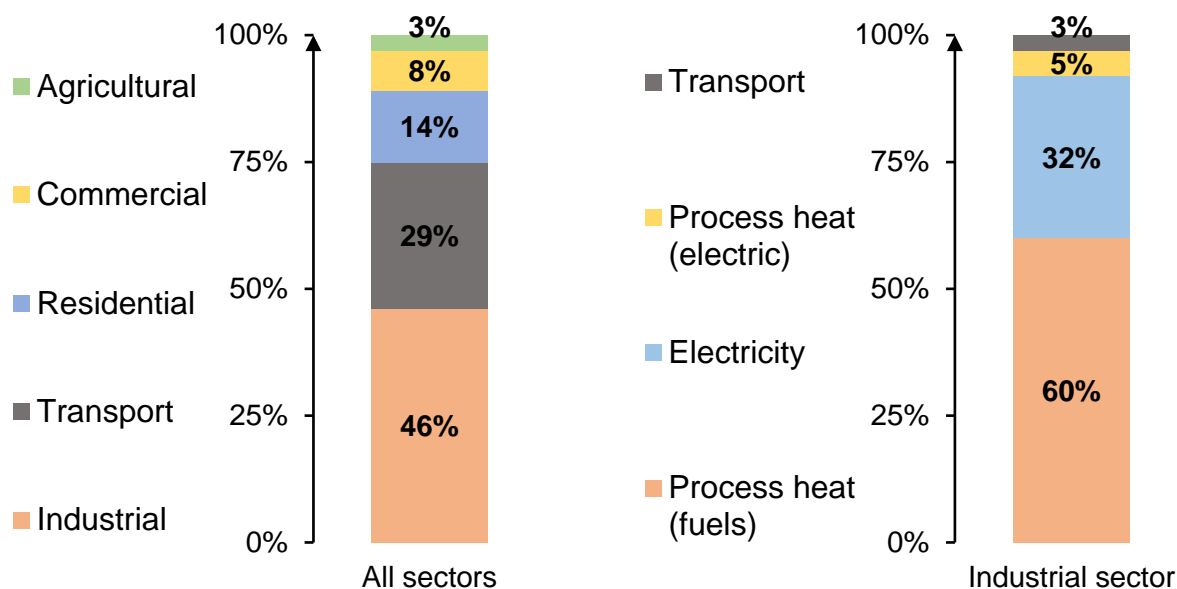
Figure 2-1: [Left] Energy demand per economic sector in South Africa, data processed by Joubert et al. (2016). [Right] Energy demand within the South African industrial sector, data processed by Hess (2016b). Data for 2006 from SATIM (2013) Appendices V3.2.

From the data it is clear that significant amounts of energy are required to provide sufficient process heat to South African industry and, currently, most of this energy is supplied by fossil fuel based

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sources. Meaningful reductions in the process heat demand by way of energy efficiency measures, or adopting renewable energy sources, could have a substantial impact on the amount of energy required from conventional energy sources in the country. Renewable energy sources would have to compete with conventional sources on an economic basis, however, even in cases of less expensive fossil fuel alternatives they might have other advantages to motivate their use.

2.1.2 Conventional energy sources

Fuels used for heating purposes

Coal is the most commonly used fossil fuel in South Africa, owing to its low cost and abundant resources. South African coal is easily accessible, resulting in low production costs (Pegels, 2010). South Africa is rich in coal resources (DoE, 2015) and at the end of 2016 had proven reserves of 9 893 million tonnes, a supply of 39 years at the 2016 production rate (BP, 2017), while other estimates state a resource of 53 billion tonnes and almost 200 year supply (Eskom, 2016).

Coal is the primary fuel used for heating purposes in multiple economic sectors. Figure 2-2 shows the shares of fuels used for heating purposes across all sectors, with a further breakdown within the industrial sector, overall, coal is used the most at 57% with oil and oil products the least at 3%. In the industrial sector 71% of the fuel used for heating purposes is coal, while it also dominates the commercial sector with a share of 87% (Joubert et al., 2016).

Coal is the most affordable fossil fuel in South Africa (Joubert et al., 2016). Competing energy sources are either scarce, or require more sophisticated processing and transport before use, resulting in much higher cost relative to coal and making them less attractive to industrial users.
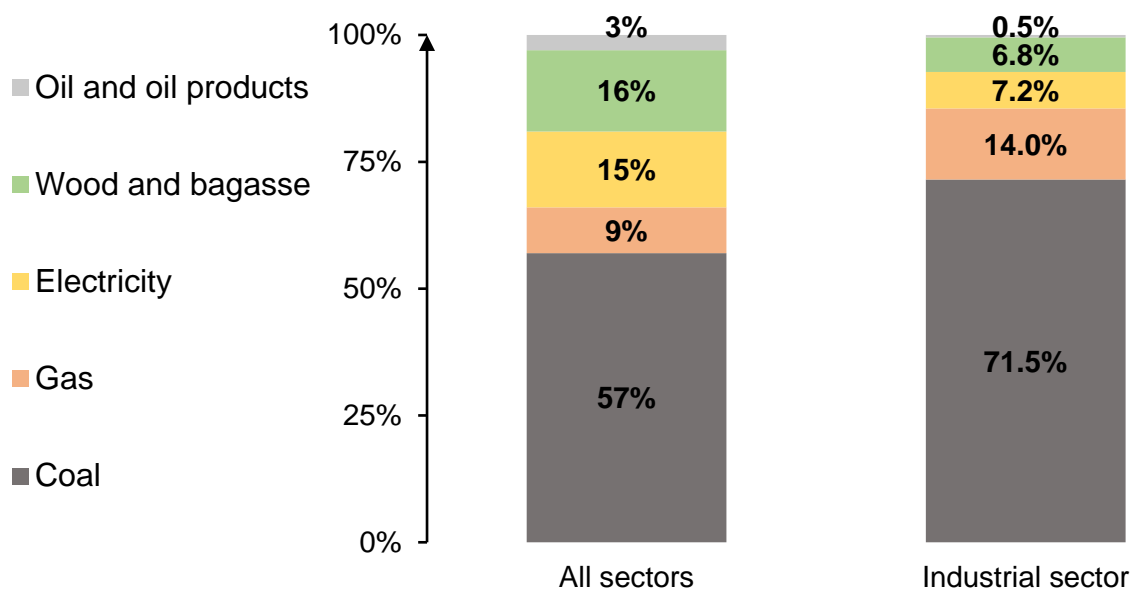
Figure 2-2: Fuels used for heating purposes across all sectors and the industrial sector specifically. Data for 2006 from SATIM (2013) Appendices V3.2, processed by Joubert et al. (2016) for all sectors and Hess (2016b) for the industrial sector.

In addition to the low cost and abundance, historical events have necessitated the use of coal that today still perpetuate its use. Historically, the need for South Africa to be independent from external energy sources due to sanctions during Apartheid resulted in both fuel and electricity being produced from coal. Eskom and Sasol arose as monopolistic suppliers of electricity and fuel from coal respectively, and consequently the bulk of investment for research and development of energy systems have focused on fossil fuels (Pegels, 2010).

This legacy of coal use is still evident today in the patterns of coal use nationally: 53% is used for electricity generation by Eskom, 33% for coal-to-liquid production by Sasol and most of the remainder for metallurgical use (Eskom, 2016). In 2017 almost 84% of the country's installed electricity generation capacity used coal as energy source, this would likely increase as two coal-fired power stations are currently under construction (Eskom, 2017).

Addressing the challenges arising from conventional fuel use

The high reliance of South Africa on coal as the primary energy source means it is critical to the smooth operation and progress of the country. Two potential future challenges have been identified with this situation: firstly, the long-term cost of coal has been steadily increasing meaning that coal could become substantially more expensive in the future, and short term price fluctuations due to various factors have contributed to price instability in the coal sector, secondly, the use of fossil fuels impacts negatively on health and the environment (Kalogirou, 2004).

Gasses released from fossil fuel combustion are one of the main causes of environmental deterioration (Thirugnanasambandam et al., 2010). South Africa is emissions intensive with the highest emissions per capita on the African continent, furthermore the per capita emissions are comparable to industrialised countries (Pegels, 2010, Ziervogel et al., 2014), which is due to the exorbitant amount of coal being combusted to produce heat. The average temperature in the country has increased by more than 1.5 times the global average and extreme rainfall events are increasing in frequency, thus climate change is a real and major concern (Ziervogel et al., 2014). The country is vulnerable to the effects of climate change since it is generally water scarce, and significant portions of the population will not be able to adapt due to low income levels and reliance on subsistence agriculture (Pegels, 2010), thus climate change is starting to be seen as a threat to development in the country (Ziervogel et al., 2014). Various financial penalties are being considered to discourage the use of fossil fuels, one example: carbon taxing, is envisaged to start in 2017 (Tshehla et al., 2017).

Solar thermal (ST) technologies are promising renewable energy alternatives with the potential of replacing a significant portion of conventional fuels used for heating purposes. Including ST heat in industrial processes could reduce CO_2 emissions and aid in increasing the energy efficiency of a process (Atkins et al., 2010). Renewable technologies will become increasingly attractive as the cost

of conventional fossil fuel based sources increases. As the renewable energy market becomes larger and more established, with increased experience and competition, the cost of renewables will also decrease.

2.1.3 National solar resource

Solar based renewable energy technologies hold particular promise in South Africa, as the country's solar resources are among the best in the world (DoE, 2015, WWF, 2017). A map of the annual global horizontal irradiation on South Africa is shown in Figure 2-3. Large areas in South Africa are well suited to harness solar energy since there are large flat areas that receive high irradiation; consequently, solar as a renewable energy resource is prevalent in seven of the nine provinces in the country (DoE, 2015). Solar energy is the renewable resource with the greatest potential in South Africa (Pegels, 2010). Solar irradiation increases to the North-West of the country, with the South-East coast of the country having the lowest solar energy potential.

Solar energy is currently being underutilised in South Africa and it is expected that solar energy use will grow significantly in the future (WWF, 2017). On average the solar irradiation on South Africa is 67% higher compared to Europe (Joubert et al., 2016), thus, more energy would be produced per unit of collector area and the potential for using solar energy technology should be higher (WWF, 2017). However, despite the greater solar resource, the installed ST capacity in South Africa is significantly less than the European countries Austria and Germany: 1.2 GW$_{th}$ compared to 3.7 GW$_{th}$ and 13.2 GW$_{th}$, respectively (Weiss et al., 2017).
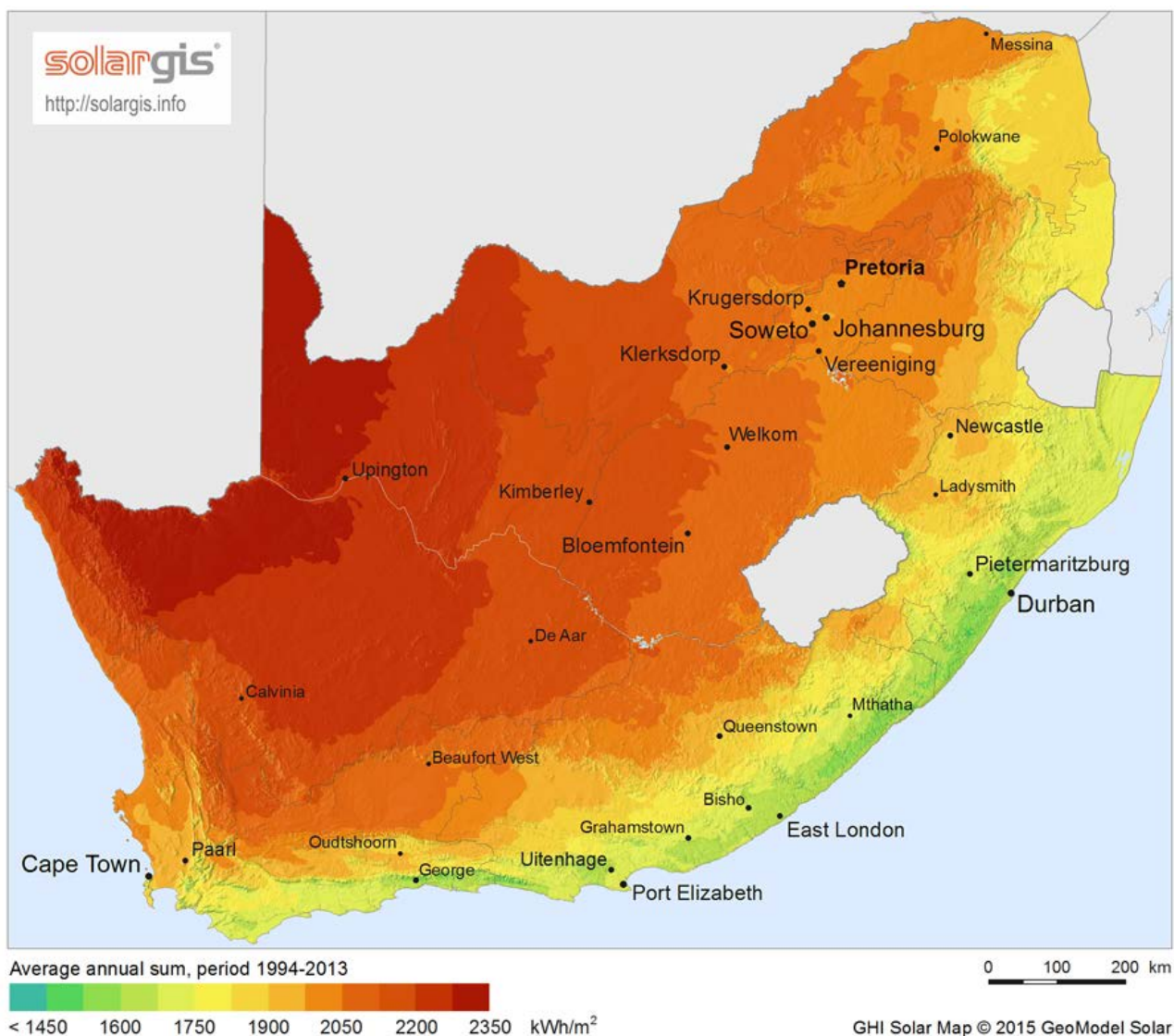
Figure 2-3: Global horizontal irradiation on South Africa. Source: SolarGIS (2015)

2.2 Solar thermal renewable energy

2.2.1 Solar thermal collector technology

Operating principle, efficiency and losses

Large amounts of solar energy radiate through the atmosphere and reaches the surface of the earth, however, due to the intermittent availability and low energy density the radiation cannot provide continuous energy supply, necessitating collection and storage methods. Solar thermal collectors (STC's) are a relatively mature branch of technology and have high efficiencies if operated in the appropriate temperature range (Weiss, 2016).

A solar collector is a specialised heat exchanger, which in ST applications converts solar irradiation on a surface into thermal energy by heating a fluid (Tian and Zhao, 2013). Normally, the STC operates at temperatures higher than the ambient temperature and consequently heat losses to the environment occur, therefore, collector efficiency is directly related to the operating temperature

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(Horta, 2015). STC efficiency is also influenced by optical behaviour, which determines the effective amount of irradiation to reach the absorber (Horta, 2015). Optical losses are determined by the design of the collector and are a property of a specific collector, while thermal losses are determined by the relative temperature difference between collector and environment and are a function of the temperature difference. Therefore, the efficiency of a STC over a range of operating conditions can be represented on a curve known as the efficiency curve (Horta, 2015).

The efficiency curve is typically represented as a second order polynomial, see Equation 2.1 (BSI, 2013). The coefficients, c_1 [W/(m^{2}K)] and c_2 [W/(m^{2}K^2)], are the first and second order environmental heat loss coefficients and are calculated using the least squares method of statistical curve fitting on the collector testing data (BSI, 2013). Optical losses due to the design of the STC are accounted for with the constant $\eta_{optical}$ parameter in Equation 2.1. Heat losses to the environment are accounted for with the difference between the mean collector fluid temperature (T_m, average of inlet and outlet temperatures) and the ambient temperature (T_a). In addition to optical and environmental losses, the final efficiency (η_{final}) is also highly dependent on the total amount of incident solar irradiation on the collector (G_t) at any point in time. The instantaneous collector efficiency is highly dependent on the operating temperature and environmental conditions of a specific application and therefore varies significantly with time.

$$\eta_{final} = \eta_{optical} - \frac{c_1(T_m - T_a)}{G_t} - \frac{c_2(T_m - T_a)^2}{G_t} \qquad \textbf{Equation 2.1}$$

Figure 2-4 is a graph showing the effect of optical and thermal losses on the efficiency of a STC; the optical losses limit the collector to a maximum efficiency, 81% for the collector considered. Thermal losses increase as the temperature difference between the collector and ambient increase, and the efficiency curve (calculated with Equation 2.1) is a combination of the optical and thermal losses. Eventually, the thermal losses equal the heat gained from irradiation, a condition known as stagnation, where no useful energy can be obtained from the collector (Hess, 2016c).

The specific STC technology determines the shape of the curve and the temperature range where it can operate with acceptable efficiency. Therefore, even though a STC can produce heat at a high temperature, the low efficiency will discourage operating under such conditions.
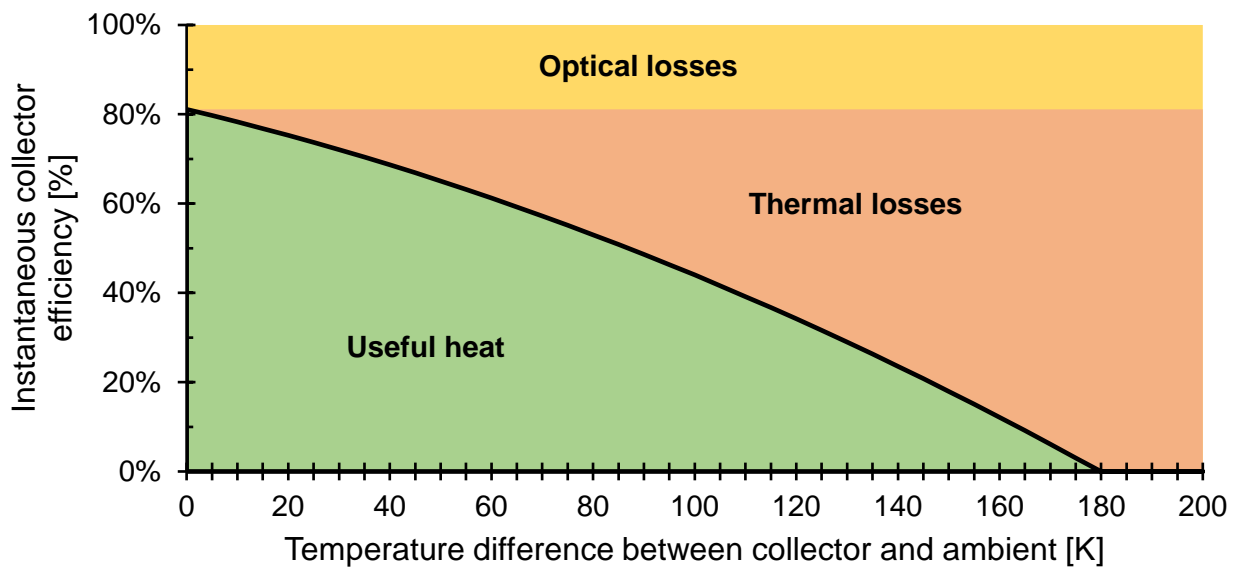
**Figure 2-4: Effect of optical and thermal losses on the instantaneous efficiency of a solar thermal collector[1].
Based on Horta (2015) and calculated for a flat plate collector with total solar irradiation of 1 000 W/m^2.**

Different solar thermal collector technologies

The role of the different STC technologies and their relation to each other can be explained with the following hypothetical discussion, based on Horta (2015). Consider the cheapest and simplest type of STC: a plastic pipe laid out in the sun with water flowing through it; irradiation will heat the pipe which will in turn heat the water. Such a system would be useful only for providing heating at a relatively low temperature, as the properties of the collector material and high environmental losses will limit this collector to very low temperatures. Improvements to collector performance can be achieved by altering the collector material and design. By using a metal, ideally one with a high heat transfer coefficient such as copper, the collector can operate at higher temperatures. To intercept as much radiation as possible the collector should have a large surface area facing the general direction of the sun. To absorb a greater fraction of the solar irradiation this metal can be painted. Heat losses via conduction, convection and radiation from the absorber can be reduced by insulating all sides not directly facing the sun. To further reduce heat losses a transparent cover can be applied to reduce convective losses from the surface intercepting solar irradiation. At this point the collector is known as a flat plate collector (FPC), a cross section of a typical FPC is shown in Figure 2-5.

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[1] The efficiency curve for a flat plate collector is shown, constructed as prescribed in ISO 9806:2013. Beam irradiation of 850 W/m^2, diffuse irradiation of 150 W/m^2, incidence angle of 0° and incidence angle modifiers set to 1. The flat plate collector for which the curve was calculated had c_1 = 2.71 W/(m^{2}K) and c_2 = 0.01 W/(m^{2}K^2).

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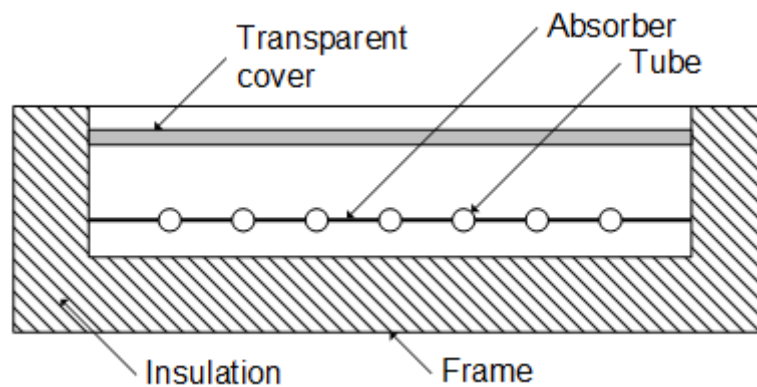
Figure 2-5: Cross section of flat plate collector. Based on Weiss and Rommel (2008) and Hess (2016a).

Losses due to convection between the absorber surface and the transparent cover can still be significant, furthermore, this is reduced by removing the air present and creating a vacuum. Some developments have aimed at creating advanced FPC's that operate with a vacuum between the cover and absorber (Weiss and Rommel, 2008), however, a much simpler design is to place the absorber in a glass tube that contains the vacuum. Practically, this requires a much smaller absorber, however multiple vacuum pipes can be placed in parallel to create a single collector. This is known as an evacuated tube collector (ETC).

The dilute nature of solar irradiation requires it to be intercepted over a large area to obtain a useful output, however, this increases thermal losses as they are directly related to the absorber area. This challenge can be overcome by using a separate reflector that concentrates radiation onto a much smaller absorber, thereby reducing heat losses. A parabola is a useful shape for this purpose, as any line parallel to its axis that reflects from the surface will be focused on a central point. Multiple parabolas are combined to form a reflector with a low concentration ratio, this can be installed underneath evacuated tubes to form a compound parabolic concentrator (CPC). Significantly higher concentration can be obtained by using a parabolic shaped reflector that concentrates solar irradiation on its central focal line, where the absorber is located. This is known as a parabolic trough concentrator (PTC). The Fresnel principle can be used to divide a parabolic reflector into segments placed on a horizontal plane concentrating radiation on a central receiver, this is known as a linear Fresnel reflector (LFR).

As the ability of a collector to concentrate solar irradiation increases, the conditions under which it can do so becomes more restrictive, ultimately requiring the tracking of the sun during the day to enable maximum utilisation of the available direct radiation. Therefore, STC's are classified mainly as stationary or tracking (Horta, 2015).

The STC technology to be used should be selected based on the required operating temperature which is determined by the specific application. Table 2.1 is a summary of the most common commercially

available STC's. Although subsequent collectors improve on the inefficiencies of the FPC, this does not mean that they are to be preferred above FPC's in all applications as each technology is well suited to a specific application due to its cost and thermal performance.

Table 2.1: Temperature ranges at which commercially available solar thermal collector technologies typically operate at acceptable efficiencies.

Movement	Collector	Indicative working temperature range	Reference
Stationary	FPC	$\leq 85°C$	Weiss (2016)
	ETC	$< 120°C$	Horta (2015)
	CPC	$100°C$ to $150°C$	Horta (2015)
One-axis tracking	PTC	$120°C$ to $250°C$	Weiss (2016)
	LFR	$120°C$ to $250°C$	Weiss (2016)

Process heat collectors

A process heat collector is any STC that can be used to provide heat to an industrial process (Weiss, 2016). Most STC technologies can be used for this purpose, however, the larger scale and more demanding environment of the industrial sector require collectors that are better suited to these applications with regards to (Horta, 2015):

- Modularity: these collectors must enable large collector field construction with fast installation and repair times.

- Robustness and safety: these collectors must endure the industrial environment and operate safely under extreme conditions, for example during stagnation.

- Operation and maintenance requirements: the existing technical personnel of the facility where it is constructed must be able to operate and maintain the system without specialised knowledge or training.

- Integration into existing processes: the ST system must be compatible with the existing system and require very limited adaption of the existing facilities.

2.2.2 Solar heat for industrial processes

Industrial solar heat installations

Solar heat for industrial processes (SHIP) describes the application of ST technology to provide heat specifically to an industrial process (Epp and Oropeza, 2017). A typical SHIP installation consists of a STC field, some form of heat exchanger to transfer energy to the process and usually a heat storage device (ESTIF, 2015), a typical hydraulic system concept is shown in Figure 2-6. Under normal operating conditions, the fluid circulating in the collector loop is heated due to solar irradiation, the fluid usually consists of water combined with ethylene glycol to prevent freezing if the ambient

temperature falls below the freezing temperature of water, however, other fluids such as thermal oils may be used for high temperature applications. The collector loop heats up the storage volume through a heat exchanger, heat storage devices enable the system to provide energy even at times when solar irradiation is not available, the most common being a thermally stratified tank. Water at the appropriate (or highest) temperature is withdrawn from the storage and integrated within the process. The heat sink may be a process unit, in which case a heat exchanger is used to transfer heat, however, if the heated water is consumed (as with water preheating applications), there is no return stream to storage and a make-up stream replenishes the storage volume. If the heat sink is a process unit that requires a fixed amount of heat, the conventional process heat source is used as auxiliary supply to provide the additional required heat (Hess, 2016a).
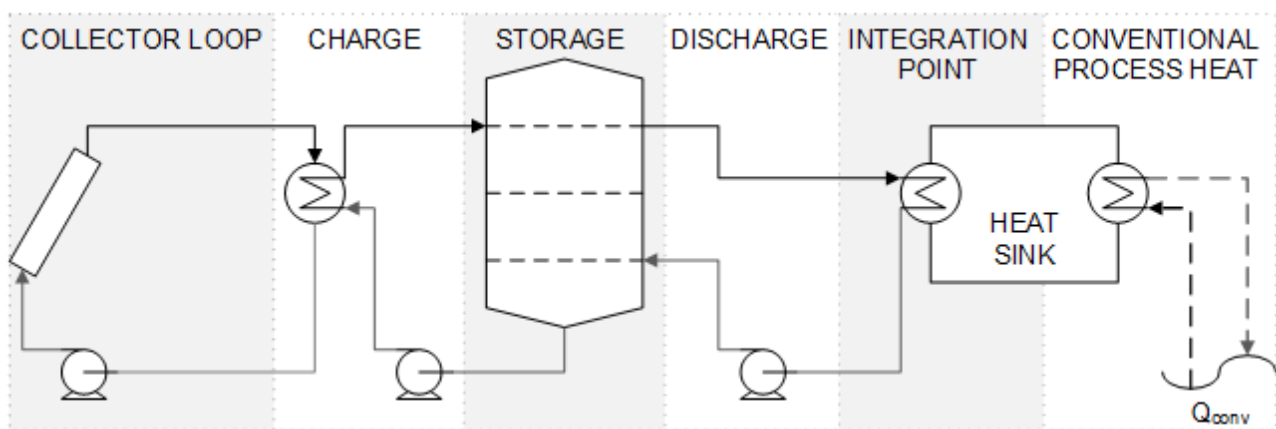
Figure 2-6: Hydraulic system concept for solar process heat integration. Based on Muster et al. (2015)

ST applications are categorised by temperature level: low, medium and high; determined by the temperature at which the specific process operates. Currently, there does not appear to be consensus about the definition of each temperature level, with the temperature authors consider to be low varying from 90°C to 250°C (ESTIF, 2015). In this project the definition given by Horta (2015) will be used as it appears to be the most widely accepted in the ST community. Low temperature being below 100°C, medium between 100°C and 250°C, and high temperature applications being above 250°C. Solar process heat applications are limited to the low and medium temperature ranges as the additional costs, safety concerns and operational complexity of higher temperature applications would make it unsuitable to most industrial users (Horta, 2015).

Favourable industries and process conditions

The technical feasibility of a solar process heat installation is mostly determined by the heat demand profile and the temperature at which heat is required. Generally, the following conditions are beneficial to the success of a solar process heat installation (Hess, 2016a):

- The conventional energy source is expensive
- The process has a constant heat demand for at least three quarters of the year

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- A process return temperature below 50°C

- Heat recovery is not technically or economically feasible

- Future changes to the facility will not affect the solar installation

Some industrial processes that operate at temperature levels suitable to solar heat are: sterilising, pasteurising, washing and cleaning (Kalogirou, 2003). The food processing industry contains several of these processes and correspondingly has a high SHIP potential. Appendix A contains more information on industrial processes with temperature ranges suited to solar heat.

2.2.3 The state of the solar thermal industry in Southern Africa

Market development and current applications

At the end of 2015 ST systems in Sub-Sahara Africa[1] accounted for about 0.3% of the total global ST installed capacity (Weiss et al., 2017), the same as the previous two years (Mauthner et al., 2016, Mauthner et al., 2015). Despite the share of global ST capacity remaining constant between 2013 and 2015, the Sub-Sahara Africa market has performed well in terms of newly installed capacity during this time. Relative to the newly installed capacity of the previous year, this market performed better than the global market during this time, showing growth during most years while that of the global market has mostly declined (Weiss et al., 2017).

South Africa is the largest contributor to the Sub-Sahara Africa ST market, with most of the installed capacity providing heat for domestic applications. At the end of 2015 approximately 90%, or $1.78{\times}10^6$ m^2, of the installed ST collector area in Sub-Sahara Africa was registered to South Africa (Weiss et al., 2017). In 2016 there were at least 89 recorded large scale[2] ST installations in South Africa (Joubert et al., 2016). Figure 2-7 is a graph showing the specific applications of the recorded systems; the largest application was for domestic hot water at 69% of the installed area, with only a small fraction of 7% providing process heat. Kalogirou (2003) reported a similar finding for Cyprus in 2003, where domestic hot water solar systems were very successful, however, there were no industrial process heat applications to be found.

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[1]At the end of 2015 Sub-Sahara Africa consisted of: Botswana, Burkina Faso, Ghana, Lesotho, Mauritius, Mozambique, Namibia, Senegal, South Africa, Zimbabwe.
[2]Systems with a gross collector area greater than 10 m^2.
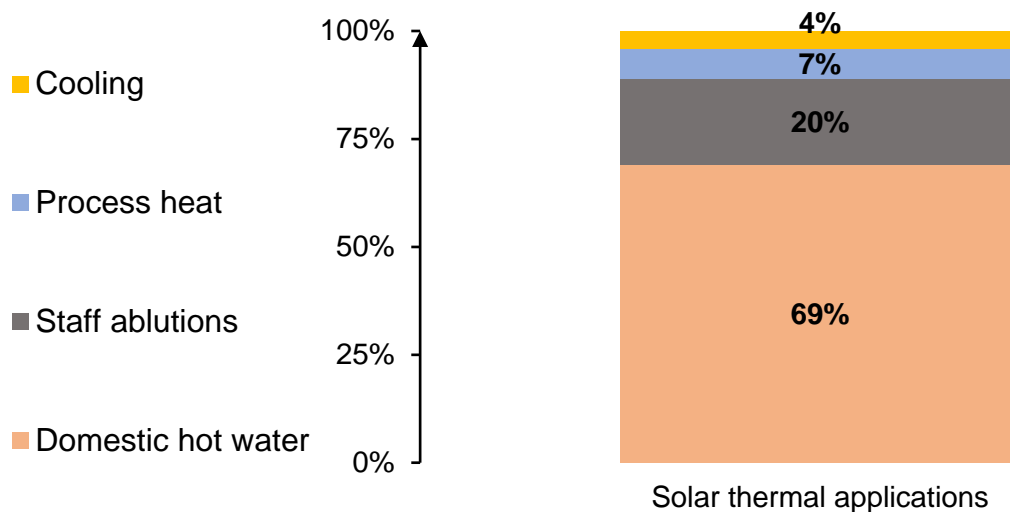
Figure 2-7: Applications of large scale (gross collector area > 10 m²) solar thermal installations in South Africa. Domestic hot water refers to all types of systems where hot water is prepared for a facility with permanent residents. Data from Joubert et al. (2016).

South Africa most likely has a very large proportion of all ST applications being swimming pool heating, however, these systems are typically small and was not included in the study by Joubert et al. (2016). Of the total installed capacity in operation in 2015 for Sub-Sahara Africa, 46% was for domestic hot water systems in single family houses, 2% large domestic hot water systems and 52% for swimming pool heating (Weiss et al., 2017). Since South Africa has the largest proportion of installed area in Sub-Sahara Africa, it is likely that the shares of applications for Sub-Sahara Africa will be most representative of South Africa. Differences in reported ST applications between Sub-Sahara Africa and South Africa are most likely due to the types of STC included in the individual studies. Joubert et al. (2016) reported the applications for large scale systems only, while Weiss et al. (2017) only considered unglazed collectors, glazed FPC's and ETC's.

Implementation of solar process heat by the South African industry

The use of SHIP is still very limited in South Africa, despite having a large and successful domestic-hot-water solar industry. The uptake of SHIP has been slow globally, despite low-temperature ST being a viable technology (Atkins et al., 2010, Lampreia, 2014).

In the South African context, the high capital cost of SHIP installations, combined with the low cost of traditional energy sources, results in long payback periods, which is unacceptable to industry (Atkins et al., 2010, Epp and Oropeza, 2017). Further, the intermittent nature and low intensity of solar irradiation (Atkins et al., 2010) coupled with the relatively high unit cost of ST energy compared to that of fossil fuels (Pegels, 2010), are additional barriers to the implementation of solar process heat. However, with increasing fossil fuel cost, stricter environmental regulations and various penalties for noncompliance ST is becoming more attractive (WWF, 2017).

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The ST market for domestic water heating in South Africa has grown considerably and is more established than the industrial heating market. The government rebate system for domestic solar geyser installations is a possible reason for the success of the domestic water heating systems (Pegels, 2010). WWF (2017) identified six factors specific to South Africa, that would aid in advancing the uptake of ST technology for use in industrial processes, these are:

- Increasing costs of electricity and fuels
- The role of the technology in decreasing greenhouse gas emissions
- Incentives promoting energy efficiency
- The cost-effectiveness of new builds compared to retrofitting of existing facilities
- Contract agreements that allow the purchase of energy from energy service companies without paying for the renewable energy installation
- The Southern African ST Training and Demonstration Initiative (SOLTRAIN)

The market for renewable energy technologies in South Africa is still in its infancy and thus has significant risk and volatility (Pegels, 2010). Uncertainty about the expected costs and benefits of a ST system hinders industry from committing to this technology (Kalogirou, 2003), consequently, there are a limited number of SHIP installations with little visibility and low awareness of the technology among industrial installations (Epp and Oropeza, 2017).

Building industry awareness and establishing a credible track record for industrial applications of renewable energy solutions are crucial to the successful implementation and uptake of the technology. Ultimately, the main factor contributing to the embracement of ST technology could be the opinion of industry formed by past experiences and not the performance of the system or the suitability of the design (Cohen et al., 2014). The support of government and intergovernmental organisations helps to increase the tempo of renewable energy implementation.

Policy and guidelines for solar energy in South Africa

From a policy and regulatory perspective, the South African industry is well positioned to implement solar energy solutions. The Solar Energy Technology Roadmap (SETRM), developed between 2010 and 2015, is a guide for the development and deployment of solar energy technologies in South Africa, while considering the relevant policy context and national incentives. SETRM focusses on three sectors: concentrating solar power, solar photovoltaic and ST technologies (DoE, 2015).

For the ST industry specifically, the South African Solar Thermal Technology Roadmap (SA-STTRM) is a guide for solar heating and cooling in South Africa with special focus on solar water heating. The SA-STTRM estimates that 4 GW of solar water heating can be installed in South Africa by 2050 (DoE, 2015). For solar heating and cooling applications in the industrial, commercial and

multi-family residential sectors a projected growth of 45% per year is required to reach the goal of approximately 4×10^6 m^2 of installed area by 2030 from roughly 10 600 m^2 in 2014 (SOLTRAIN, 2015).

There are numerous international programs that promote the use of ST technology. Solar Payback (Epp and Oropeza, 2017) is a program between Brazil, Mexico, India and South Africa that promotes SHIP and attracts investors by raising awareness of its technical and economic potential.

2.3 Fishmeal and fish oil

2.3.1 Properties, advantages and uses

Raw materials and physical properties

Fishmeal is a stable, high protein concentrate available as a powder, and used in the animal feed industry as a source of high quality protein (Barlow and Windsor, 1984). It is produced from small pelagic fish or fish processing by-products; by producing fishmeal, the volume of materials needing transport is greatly reduced, and the product lifetime significantly increased since fishmeal will not support microbial growth causing spoilage (IFFO, 2016, Windsor, 2001). Fishmeal can be stored for several years while mostly maintaining the nutritional value thereof (Windsor, 2001).

Fishmeal is the solid product obtained after removing most of the water and a fraction of the oil within raw fish materials (Windsor, 2001). Therefore, the composition of the fishmeal product reflects that of the starting material, and the quality is highly dependent on the starting materials (IFFO, 2016). The protein content of fish used to produce fishmeal is approximately constant at 16% (Barlow and Windsor, 1984), as physiological processes in the fish body maintains the combined portion of oil and water at relatively constant levels irrespective of the fish species. Standard fishmeal typically contains 64% to 67% crude protein and up to 12% oil, while special fishmeal may have a protein content up to 72% (IFFO, 2016).

Fishmeal production increases food security by converting harvestable fish, which are not consumed by humans, into animal feeds which are then used to farm other animals which are directly consumed. The fish species commonly used for fishmeal production are mostly not desired, and in some cases unfit, for human consumption, and are small, bony, fast-growing fish with short lifespans and high oil content (IFFO, 2016). The majority of these fish are found in the upper layers of the sea and are therefore known as pelagic (Pike and Jackson, 2010).

The proportion of fishmeal being produced from by-products is increasing globally, consequently, due to the stagnant production of fishmeal, the amount being produced from wild caught fish is decreasing. The use of by-products from fisheries and fish processing to produce fishmeal is

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increasing; in 2012: 35% of fishmeal was made from by-products (FAO, 2014). Despite an increase in the amount of oily fish consumed by humans the production of fish oil is expected to remain relatively constant, due to an increase in fish by-products being used for fish oil production (Pike and Jackson, 2010).

Fishmeal as a compound animal-feed ingredient

The main use of fishmeal is as an animal feed ingredient. Standard fishmeal is used in feeds for poultry, ruminants and omnivorous fish, the more expensive special fishmeal is used for more sensitive species like carnivorous fish, crustaceans and swine (IFFO, 2016). Fishmeal in animal diets is a good source of protein, essential amino acids, energy, minerals, vitamins and essential fatty acids (Barlow and Windsor, 1984). The inclusion of fishmeal in compound animal feeds makes it more palatable and improves nutrient utilisation, which helps to maintain a healthy immune system (Miles and Chapman, 2005). Fishmeal further not only provides high levels of protein, but fish proteins are also known to contain high levels of essential amino acids which cannot be synthesized by animals and therefore need to be ingested as part of their diet (Cho and Kim, 2011). The essential amino acids are also more utilizable in fishmeal than other meals (Windsor, 2001).

The majority of fishmeal produced globally is used in compound feeds for aquaculture, which are in greater demand than ever before. The contribution of aquaculture to human fish consumption globally has increased from 5% to 49% between the 1960's and 2012, and the global average per capita fish consumption has increased from 9.9 kg to 19.2 kg in the same period (FAO, 2014). The demand for compound feeds for the aquaculture industry, and consequently the demand for fishmeal, has increased considerably due to the increased fish consumption and the greater portion of fish supplied by aquaculture.

The constant supply of fishmeal over the past few years and significant increase in demand by the aquaculture industry has resulted in fishmeal being considered a strategic ingredient. It is to be used sparingly and only during periods in the animal lifecycle when it will have the biggest effect.

The uses of fish oil

The main use of fish oil is in the aquaculture industry, as a part of the diet of carnivorous fish (Pike and Jackson, 2010). High quality fish oils may also be used in the pharmaceuticals industry (Windsor, 2001). It also has diverse other uses: as a carrier for pesticides, in paints and varnishes, in the leather industry and as soaps and greases (Pike and Jackson, 2010, Windsor, 2001, FAO, 1986).

Fish oils have a high concentration of long chain, polyunsaturated fatty acids, especially omega-3 fatty acids, making it unique in comparison to other fats obtained from plants and animals (Barlow and Windsor, 1984). The inclusion of these fatty acids in the human diet is beneficial for health,

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especially cardiovascular health, as well as neurological development and mental health (Pike and Jackson, 2010).

2.3.2 The fishmeal and fish oil production process

Fishmeal manufacture is a well-established process and generally comprises of sequentially cooking, pressing, drying and milling the fish raw materials. The different operations in this process may be simple, however, considerable skill and experience is required to separate the components efficiently and at a low cost. (Windsor, 2001). The most common commercial fishmeal production process is known as the wet-pressing method (IFFO, 2016).

The various cooking, separation and drying stages of the wet-pressing method can be viewed as chemical engineering unit operations. A general block flow diagram (BFD) representative of most wet-pressing fishmeal production processes is shown in Figure 2-8. Fishmeal production methods have been well established for a number of years and the process has generally remained the same.

The unit operations that constitute the wet-pressing fishmeal production process is briefly described below and should be read in combination with the BFD shown in Figure 2-8. The unit operations are described in more detail in Chapter Three.

i. Cooking of the fish at 85°C to 100°C coagulates the protein, ruptures fat deposits and liberates oil and water.

ii. Straining separates some of the oil and water liberated during cooking from the solids.

iii. Pressing of the solids in a screw press separates the liquor from the solids, solids exiting the press are known as the press cake, which are then sent to the dryers as a press cake with approximately 50% water content.

iv. Liquor:

    a. Decanting removes more of the suspended solids, which joins the press cake as a sludge, known as grax.

    b. Centrifugation of the decanted liquor separates the oil from the aqueous phase. The aqueous phase has a high viscosity and tends to be sticky, therefore, it is called stickwater.

    c. Concentration of the stickwater by evaporating a fraction of the water creates the stickwater concentrate, which joins the press cake entering the dryer.

    d. Polishing of the fish oil ensures the correct quality for the intended purpose.

v. Drying of the press cake, decanted solids and stickwater concentrate indirectly with steam at 800 (170°C) to 1 000 kPa (180°C), or directly with heated air (at approximately 500°C), raises

the temperature to approximately 90°C; this forms a stable meal with a water content of roughly 10%.

vi.   Milling ensures a final product of similar particle size and improves product handling.

vii.   Storage of the fishmeal can be in either 25 kg, one-ton bags or bulk warehouses.

To describe the changes the fish raw material undergoes as it is processed to fishmeal, a mass balance is shown on Figure 2-8, with generic composition values for typical industry performance shown (Windsor, 2001). The fish raw materials entering the process has a solids content of 18% (1). Straining and pressing creates a stream which is mostly water with a low solids content of 6% (3 and 5) and another stream with a much higher solids content of 44% (8). Decanting the liquor (3 and 5) removes some of the suspended solids and the exiting stream (6) has a slightly lower solids content of 5%. Centrifugation separates the incoming stream (6) into the fish oil (7) and aqueous stickwater streams (10). Concentration of the stickwater greatly increases the solids content from 6% (10) to 33% (12). The combined stream entering the dryers (8, 9 and 12) has a solids content of 41% which is increased to 85% and a final water content of 9% (13).

1	Raw fish
2	Cooked fish
3	Liquids
4	Solids
5	Liquids
6	Liquids
7	Fish-oil
8	Press cake
9	Decanter grax
10	Stickwater
11	Polished fish-oil
12	Stickwater concentrate
13	Dried fishmeal
14	Finished fishmeal

Stream(s)	Composition (Mass %)		
	Solids	Water	Oil
1	18	70	12
3 and 5	6	78	16
6	5	78	17
7	0	0	100
8	44	53	3
9	33	67	0
10	6	94	0
12	33	65	2
8, 9 and 12	41	57	3
13	85	9	6
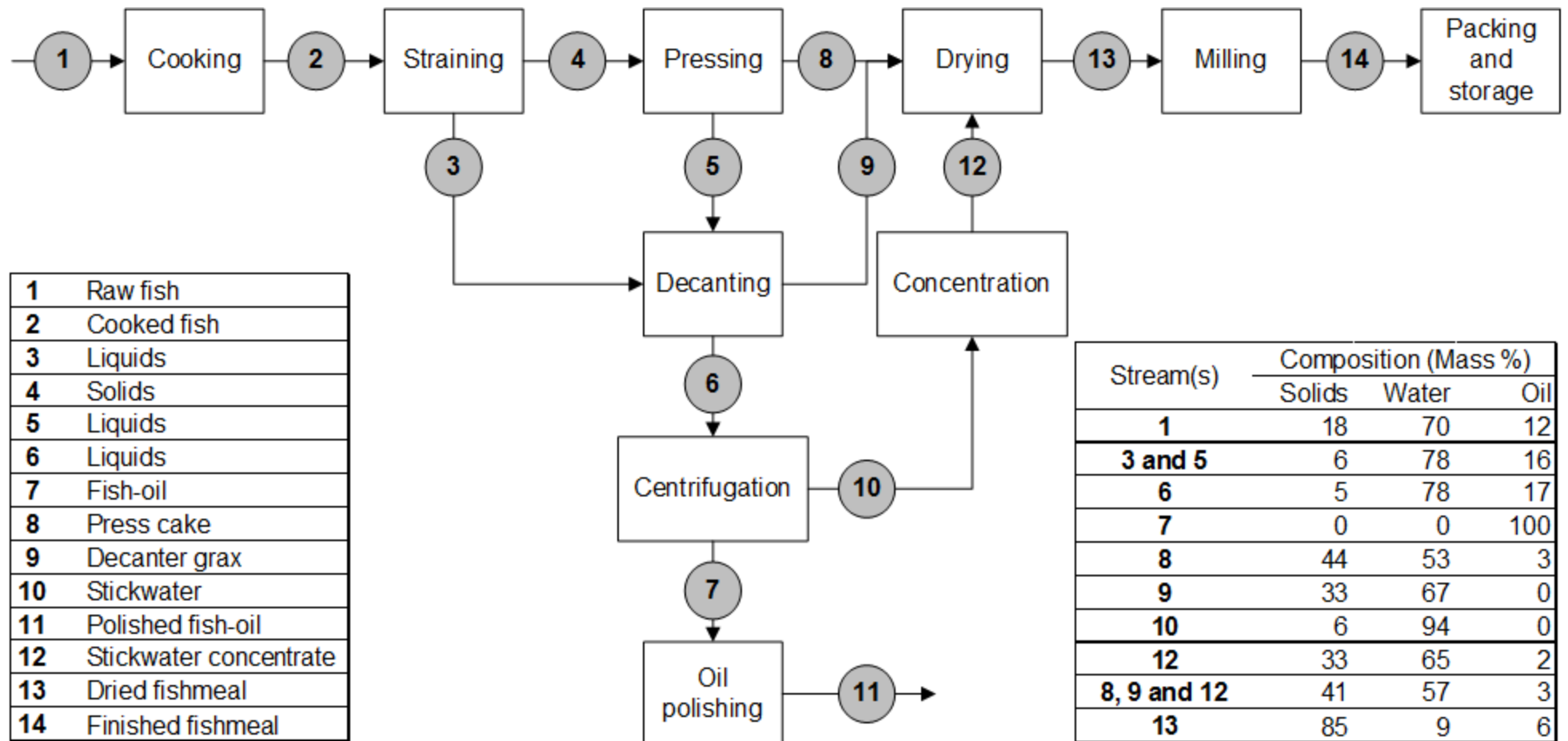
Figure 2-8: Block flow diagram and mass balance of wet-pressing fishmeal and fish oil production process, based on IFFO (2016), Windsor (2001), FAO (1986), and Barlow and Windsor (1984). Compositional values are for a generic process based on typical performance encountered in the global fishmeal industry.

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2.3.3 Fishmeal production in South Africa

The fishmeal and fish oil (FMFO) production industry is important to the South African economy and is a mature industry in the country (de Koning, 2005). It contributes substantially to the GDP and provides employment to a significant number of people, either directly in the production facilities or indirectly through the local pelagic fishery (Hara et al., 2008). The pelagic fishery provides employment to approximately 5300 people and supports 8 FMFO plants and ancillary facilities (Hara et al., 2008).

In South Africa FMFO is produced from the pelagic species: anchovy and round herring, as well as by-products from processing fish destined for human consumption (Hara et al., 2008). The FMFO industry is dependent on fisheries to provide the required raw materials, whereas the fisheries are limited by total allowable catch quotas and fishing seasons. The individual quotas have been known to vary significantly between years, especially in the pelagic fishery (Hara et al., 2008), but have mostly decreased with time (Japp, 2001), due to a greater number of licenses allocated and stricter environmental regulations. The pelagic fishery is the largest in terms of volumes caught and in 2005 accounted for more than 70% of the total catch (Hara et al., 2008). The dependence of FMFO factories on the availability of fish and the amounts caught result in fluctuations during the production year, at times there may be a lack of raw material and at other times abundance (Windsor, 2001).

Over the previous twenty years the global fishmeal market has become less stable with the magnitude and frequency of short-term price fluctuations increasing. Figure 2-9 is a plot of the price of fishmeal from June 1997 to June 2017, most of the fishmeal produced in South Africa is exported and thus the price in United Stated Dollar (USD) and South African Rand is shown. Until mid-2006 the global fishmeal market was reasonably stable, the price of fishmeal was below 1 000 USD and fluctuations were relatively low. From 2007 the price of fishmeal has mostly increased with greater fluctuations occurring, further drastic changes in the short-term price of fishmeal were more frequent, for example the sharp increase between June 2009 and June 2010. South African producers face even more precarious prices due to the exchange rate between the Rand and USD. Comparing the prices of June 2017 to June 1997, the average increase in Rand has been about 6.2% per year above inflation of 6%, despite the decrease in price from June 2016.
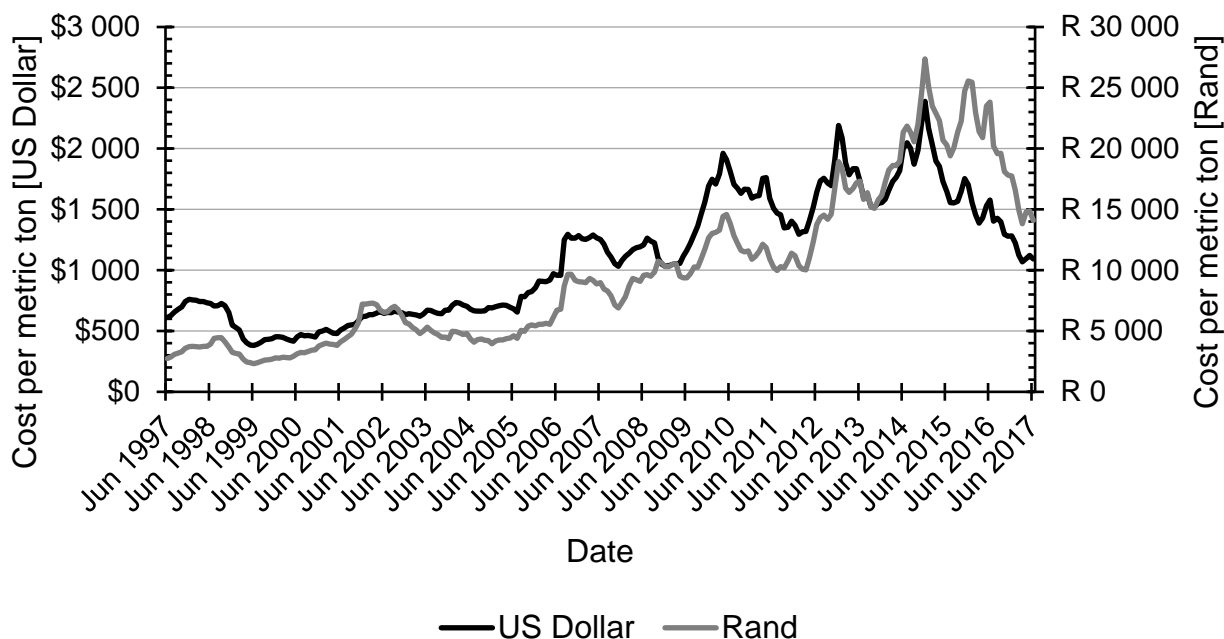
Figure 2-9: Selling price of fishmeal over the past 20 years, presented in United States Dollar (GEM Commodities World Bank Group, 2017c) and South African Rand (GEM Commodities World Bank Group, 2017b), values shown are not adjusted for inflation.

2.4 Concluding remarks

The future sustainability of the South African industry is challenged due to its reliance on fossil fuels for heating purposes. Conventional energy sources may no longer be the most attractive option from an economic perspective in the future, since the costs of fossil fuels are expected to increase indefinitely in the long term. A more urgent concern is the need to reduce fossil fuel combustion gas emissions due to their detrimental effect on the environment.

ST renewable energy has the potential to reduce the amount of fossil fuels consumed, by substituting a significant portion of the low to medium temperature heat that is currently supplied by combusting fossil fuels, with heat obtained from solar radiation. Solar energy can aid in stabilising energy costs since it uses freely available solar radiation.

The technical and economic performance of a ST heating system is unique to each application and production facility, as the efficiency of the system to convert solar radiation into useful heat depends on the specific temperatures involved, the production schedule and the instantaneous irradiation at the location. Therefore, detailed knowledge of the operations of the facility and a thorough understanding of the process unit-operations and their interactions are required to design a ST system for a facility.

The FMFO industry in South Africa is in a favourable position to benefit from ST heating, provided feasible opportunities for solar heat integration can be identified and implemented. These FMFO factories typically have considerable process heat demand that is currently provided by fossil fuels,

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while also being located in areas of relatively high solar irradiation. The cash flow of FMFO factories is volatile due to fluctuations on both the operational cost side (fluctuating fossil fuel costs), and on the revenue side (fluctuating product prices). This cash flow volatility can be reduced on the operational cost side by integrating solar heat, which results in more constant energy costs since the solar radiation source remains free.

The energy requirements may differ vastly between different FMFO factories, even though the production method is well established and this is a mature industry in South Africa. This is due to differing production schedules, type of raw material utilised, operating conditions, process topology and equipment conditions. Therefore, the opportunities for solar heat integration will differ between different factories and will have to be investigated for each production facility, based on their environmental and operating conditions. However, a solar heat integration study performed for a factory that uses the wet-pressing method with high throughput rates may provide insights into the FMFO industry, since the unit-operations used are generally the same for factories based on this method.

3 Describing the production process and resource requirements of two typical South African fishmeal plants

The definition and characterisation of the base case processes for two South African fishmeal factories is presented in this chapter. The wet-pressing fishmeal production method is described in detail, together with a short description of alternative production methods and recent research on the energy consumption of fishmeal factories. The motivation for studying the specific factories is given, and the methodology used to define and characterise the base case processes is presented. The development of the base case process simulations for the two factories is presented. Lastly, the two factories are compared and the results of the base case processes compared to available literature values.

3.1 General process description of fishmeal production methods

3.1.1 Wet-pressing method for fishmeal and fish oil production

Most FMFO producers globally use the wet-pressing method (IFFO, 2016), which is characterised by separating liquids from solids by mechanical pressing after cooking of the raw material. Regardless of the specific process used, almost all industrial fishmeal production uses specialised equipment intended for that task (Windsor, 2001) A detailed discussion of each stage in the wet-pressing method follows.

Cooking

Cooking of the raw material is the first step in the fishmeal production process. This step coagulates the proteins in the raw material and releases the physio-chemically bound water and oil, which is necessary since mechanical pressure alone will only release a small amount of liquid (Windsor, 2001). Cooking is critical to the success of the process as incomplete cooking will not satisfactorily release the liquor while overcooking will render the material too soft to be pressed (Windsor, 2001), resulting in a high concentration of suspended solids in the stickwater (FAO, 1986).

The cooking operation is also critical in preserving the material as it disables microbial and enzymatic activity, thereby reducing spoilage of the fish (IFFO, 2016). Nygaard (2010) recommended cooking for 20 minutes at 70°C for wild fish and 76°C for fish from aquaculture, as the minimum conditions to reduce microbial activity to acceptable levels in Norwegian fishmeal production processes. These values should generally apply to all facilities processing whole fish from the sea and aquaculture fish or products that did not show signs of disease.

The cooking process heats the raw material to temperatures in the range of 85°C to 100°C (IFFO, 2016, FAO, 1986) within a short amount of time. It has been shown that the fat deposits are liberated at 50°C and that the protein coagulates quickly at a temperature of 75°C, thus there should be no reason for heating of the fish materials above 75°C (de Koning, 2005). In practice, however, the higher temperatures of approximately 100°C are preferred as this is believed to result in better operation of fishmeal production plants, this could possibly be due to downstream process requirements (Bergé, 2016).

Commercial cookers consist of a screw conveyer inside of a cylinder fitted with a steam jacket (Windsor, 2001), these process units generally use steam at 600 kPa (gauge) and 165°C (tac makina, 2017). Fish materials are indirectly heated with steam as they travel through the cooker, some cookers enable directly injecting steam into the materials being cooked (Windsor, 2001). Typical indirect cookers range in raw material processing capacity from 16 to 1600 tonnes per day. Pre-cookers may also be added before the cookers to heat the materials to approximately 50°C to 60°C, this utilises waste heat and assists in reducing scaling in the cooker (FAO, 1986).

Straining and pressing

The cooking stage is followed by straining and mechanical pressing to separate the oil and water that was liberated during cooking. A large portion of the liquids released during cooking are separated from the solids by employing a strainer conveyer, vibrating strainer or rotating strainer. Straining of liquids prior to pressing may improve the efficiency of the pressing stage by diverting large amounts of fine suspended solids (FAO, 1986). The liquid separated by screening and pressing is called press liquor and consists of water which contains dissolved and suspended solids, and oil (Windsor, 2001).

Mechanical pressing of the cooked materials further removes oil and water and reduces the water content of the cooked materials from 70% to approximately 50% (Windsor, 2001). Screw presses are commonly used, which consist of a tapered shaft located within a perforated tube that transports material along the length of the press, as the volume gradually decreases the pressure increases forcing liquid through the perforations in the tube (Windsor, 2001). The solids, known as press cake, exit via the press screw and may contain 60% to 80% of the total solids (FAO, 1986). The most efficient press is the twin screw press containing two screws mounted side by side and rotating in opposite directions (FAO, 1986).

The straining and pressing stages are not always required as they are mainly included to separate oil from the solids to enable fish oil production. Fishmeal production from white fish or their by-products does not necessitate the pressing stage due to the low oil content of the raw material, although, the straining and pressing equipment may be included to enable processing both white and oily fish. Apart from greater raw material flexibility, inclusion of this equipment can increase energy

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efficiency, since multiple effect evaporators can be used to concentrate the separated stickwater. These consume less energy to evaporate the water compared to fishmeal dryers (Windsor, 2001).

Decanting and centrifugation

The purpose of these stages are to separate the fish oil from the solids and water. Decanting removes a portion of the solids along with water, while centrifugation separates the oil from the remaining water and solids. The cost effectiveness of this step is dependent on the type of raw material processed as oil content varies significantly between different fish species. It is mostly not profitable to separate and purify the oil when raw material with less than 3% oil content is processed and thus the oil recovery step is omitted in certain facilities (FAO, 1986).

In addition to enabling the production of fish oil, these stages significantly increase the yield of fishmeal by recovering dissolved and suspended solids in the press liquor. The suspended solids and water removed from the press liquor with decanters are added back to the press cake, increasing the mass of solids. The remaining liquor is then centrifuged again to separate the oil from the aqueous phase, this requires high temperatures of 90°C to 95°C for efficient separation and reheating of the decanted liquor may be required (FAO, 1986). The aqueous phase is called stickwater since it tends to have a high viscosity compared to water (IFFO, 2016), it usually has a solids content of approximately 9% by mass (Windsor, 2001). The solids in the stickwater are recovered by evaporating a portion of the water and adding the concentrate back to the press cake.

Oil polishing

A final step known as polishing is used to remove any remaining solids in the oil and obtain the desired purity, before the fish oil is pumped to tanks for storage. Polishing of the oil obtained after press liquor decanting and centrifugation, is performed with another centrifugation step (Windsor, 2001), additionally, special filters may be used to remove impurities soluble in the oil (IFFO, 2016). Polishing is performed optimally at 95°C and is facilitated by hot water, as the process is strongly influenced by material properties which are temperature dependent (FAO, 1986).

Concentration of stickwater

The fishmeal yield is significantly increased in most cases by recovering the solids in stickwater. Stickwater makes up approximately 65% of the raw materials (FAO, 1986), depending on the equipment performance and raw material processed. Stickwater contains protein-rich solids which contributes up to 20% of the fishmeal produced if included (Windsor, 2001), therefore, it is critical that the solids are recovered and returned to the press cake.

The high water content of stickwater (up to 90%) contributes to the high energy requirement of the process, as the water needs to be evaporated from the final product. Two options exist to evaporate

the excess water in stickwater, either in the final drying stage or in a separate concentration stage using multiple-effect evaporators. Multiple-effect evaporators have proven to be more energy efficient compared to steam dryers and enable the utilisation of waste heat from the factory (FAO, 1986), thus, they are the most widely used equipment pieces in industry for recovering heat from a vapor (Myrvang et al., 2007). Table 3.1 contains the typical steam requirements and production rates required for multiple-effect evaporators. The steam requirements decrease as more evaporation stages are added as the vapour coming from previous stages are used for heating subsequent stages, however, this configuration requires high production rates to be financially viable.

Table 3.1: Typical figures for multiple-effect evaporators. Data from FAO (1986) and Myrvang et al. (2007).

Type of evaporator	Steam required per water evaporated [kg/kg]	Raw material capacity [Metric ton/day]
Single	1.1 to 1.3	Not stated
Double	0.6 to 0.65	30 to 150
Triple	0.37 to 0.45	200 to 400
Quadruple	0.2 to 0.35	500 and more

Due to the increase in yield and the higher energy efficiency compared to dryers, the use of multiple effect evaporators are encouraged and stickwater should be concentrated to as high solid-fraction as possible (FAO, 1986). Usually, stickwater is concentrated to a solids content of 30% to 50% before drying (Windsor, 2001).

Drying

Drying is the final moisture removal stage in the fishmeal production process and has a large effect on the final product quality. Insufficient drying yields a product that is susceptible to mould or bacterial growth, while excessive drying leads to a scorched product with reduced nutritional value (Windsor, 2001). To ensure microbial activity is eliminated, fishmeal should be dried to a water content below 12% (FAO, 1986).

Two types of fishmeal dryers are encountered in industry: direct air dryers and indirect steam dryers. Direct drying is quicker, and uses hot air at high temperature (up to 500°C) to dry the material as it tumbles in a cylinder. The evaporation of water from solid particles ensures temperatures of the material being dried remains below 100°C, however, if the drying process is not carefully controlled scorching may occur (Windsor, 2001). Common indirect steam dryer designs are based on a steam jacketed cylinder, a cylinder containing hollow discs heated by steam (rotary disc dryer) (Windsor, 2001) or a cylinder containing coils through which superheated steam flows (coil dryer) (IFFO, 2016).

A typical rotary disc dryer can process approximately 300 tons of raw material in 24 hours, while coil dryers have a larger capacity of up to 400 tons per day. Air dryers are capable of much larger throughput rates (FAO, 1986).

Indirect steam drying is a gentler method, however, it comes at the expense of longer drying times compared direct air dryers. With increases in cost of fuel and stricter environmental control, indirect steam drying is the preferred method. Steam dryers are also advantageous since considerably less effluent gasses are produced compared to air dryers, this results in fewer air pollution problems (FAO, 1986) and less of an odour surrounding the factory, a common complaint with the public. Indirect dryers typically have a higher efficiency since it is not necessary to heat large volumes of air prior to drying (Myrvang et al., 2007).

Presently, fishmeal is mostly dried at a temperature of 90°C since nutritional value may be effected at higher temperatures with current drying equipment (FAO, 1986, IFFO, 2016). The drying process is controlled by varying the throughput rate of the dryer or changing the steam pressure. Most steam dryers consume steam at a maximum temperature of 170°C (FAO, 1986), with a corresponding gauge pressure of approximately 700 kPa.

Milling

Fishmeal is milled to reduce the size of large lumps and bones that remain after drying, in order to ensure a uniform particle size as required by the majority of animal feed applications (Windsor, 2001). A screening stage precedes milling to remove any foreign materials present (FAO, 1986). Hammer mills are well suited to the fishmeal industry as they are capable of handling the high throughput rates and can be cleaned easily (FAO, 1986). Hammer mills reduce the fishmeal to the appropriate size by impacting it with rotating hammers, a grating ensures the particles remain until the size has been sufficiently reduced.

Packaging and storage

During storage the oils present in fishmeal can react with oxygen, this decreases the nutritional value of the product, therefore, antioxidants are added to the fishmeal. A more serious concern is that the heat generated by the oxidation reaction may cause the fishmeal to combust. Ethoxyquin is the most commonly used antioxidant (Windsor, 2001). The reactivity of different oils seems to vary and thus antioxidants are not always required. Fishmeal can be packaged in small bags of about 25 kg for local sales, or in bulk warehouses or one-ton bags for transportation and shipping (IFFO, 2016).

3.1.2 Alternative fishmeal production methods

Alternative production methods are possible, where either alternative equipment or altered production flow sheets are employed. The conventional wet-pressing method is altered due to economic considerations, for smaller factories or to produce a special product (FAO, 1986).

The raw material properties or product specification requires alternatives to the basic equipment used in the wet-pressing method, in some cases. Centrifuges are preferred over screw presses for raw materials that result in significant amounts of fine and suspended solid particles, as these pose a challenge to mechanical pressing (Windsor, 2001). Where gentle, low-temperature drying is desired to preserve the protein quality, low temperature dryers that operate with indirect hot air, or under vacuum are used (IFFO, 2016).

Alternative process flow sheets have been developed that either employ other mass and heat transfer mechanisms, or altered the wet-pressing method by removing process units. The heat transfer method is the best known alternative, where oil is added to a slurry of the raw materials and acts as a heat transfer medium (Windsor, 2001).

Fishmeal can be produced from lean fish and offal by omitting the cooking stage of the wet-pressing method (FAO, 1986). This is typically performed with two drying stages, the dryers may operate at atmospheric pressure, or the first may operate under vacuum to ease the removal of water, as with the 'Schlotterhose' method. One method produces fishmeal from oily species by drying without cooking and then removing oil from the dry fishmeal by mechanical pressure.

The wet-pressing method may be simplified by excluding the stickwater and oil separation stages, to save costs (FAO, 1986), however, this limits factories to lean fish and a low processing capacity. In this process, all the material is cooked, dried and milled without any separation occurring. With all the water being evaporated in the dryer, this process requires much more energy compared to the typical wet-pressing method.

Some applications have an extremely low tolerance for fats of marine origin, solvent extraction may be used to reduce the fat content to acceptable levels (FAO, 1986). The marine fats in the form of oils can by extracted from the dry product or the wet raw materials. Solvents like ethanol, isopropanol or hydrocarbons are used to extract oils from fishmeal, they are heated before being added to the dry meal. Wet extraction can be performed on the raw materials or the press cake. Ethylene tetrachloride is typically used, since it forms an azeotrope with water at 88°C, allowing it and the water to be boiled off, leaving the oil behind. Although, there are many objections to chlorinated solvents. Generally, after contacting the fish material the loaded solvent must be sufficiently removed from the solids, and the oil and water removed from the solvent before it can be regenerated.

3.1.3 Fishmeal classification

Fishmeal is classified according to quality, which determines its market price. Quality is mainly determined by protein content, more specifically the content of the essential amino acids methionine and lysine (FAO, 1986) and the absence of microorganisms that can cause disease in humans (Windsor, 2001). The quality of fishmeal depends on the condition of the raw materials used, for example the freshness of the material (measured as total volatile nitrogen [TVN] (IFFO, 2016)) and whether it consisted of whole fish or by-products such as fish frames and offal, and the temperature at which drying occurs (de Koning, 2005).

Higher quality fishmeal classes can cost significantly more, and the increased prices are justified in the animal feed market by the increased nutritional value of the product, with a greater proportion of the essential amino acids required by the animal present in the feed. According to Fréon et al. (2017), the three major fishmeal classifications are:

   i.   Standard, also called fair average quality (FAQ)
   ii. Prime
   iii. Super prime, also called LT (IFFO, 2016).

FAQ fishmeal is typically produced with direct air drying as the drying stage, which exposes the product to high temperature gasses (up to 500°C). Fréon et al. (2017) includes residual fishmeal, made from fish waste, with FAQ as a lower quality fishmeal. Prime and super prime fishmeal is produced by drying at lower temperatures, with indirect steam dryers (using steam at 170°C) for example, resulting in higher quality products (IFFO, 2016).

3.1.4 Reported energy requirements for fishmeal production operations

There has been very little new work done concerning the energy requirements of FMFO operations, despite the fishmeal production process being energy intensive. Two recent studies stating the energy requirements of fishmeal operations were found: one by Fréon et al. (2017), based on data collected between 2008 and 2012, and the other by Myrvang et al. (2007).

Aiming to decrease the negative impact that the Peruvian fishmeal industry has on the environment, Fréon et al. (2017) performed life cycle assessments on three fishmeal factories located in Peru, with different throughput rates and producing different classes of fishmeal. Two of the factories produced FAQ fishmeal and combined they consumed approximately 58 000 tons of raw material per year, and the other factory consumed approximately 155 500 tons of raw material per year to produce prime fishmeal.

Due to the energy intensive nature of the drying process, Myrvang et al. (2007) studied the possibility of using excess heat, in the form of low pressure steam (at 3.4 bar) from a petroleum refinery in Norway, in a fishmeal production facility. The specific Norwegian FMFO factory had daily and annual raw material capacities of 1 000 tons and 130 000 tons respectively.

The fishmeal production process energy requirements stated in these studies are valuable, since they are the most recent literature values calculated for specific, existing fishmeal factories with high throughput rates. Fréon et al. (2017) reported the energy required from fuel to convert one metric ton of raw material to fishmeal for the factories they studied, it was 1 498 MJ for prime fishmeal and 1 913 MJ to 2 406 MJ for FAQ fishmeal. The energy requirements for FAQ meal are higher since direct air dryers are typically used, which consumes more fuel. Myrvang et al. (2007) reported 1 890 MJ of energy required from fuels to process one metric ton of raw material into prime fishmeal, they also stated the corresponding steam requirement of 528 kg.

The energy required to produce fishmeal varies with the capacity of the plant, the efficiency and age of equipment, and the utilisation of waste heat (FAO, 1986), however, since similar process units are typically encountered the energy requirements fall within a range. Both prime fishmeal plants investigated by Fréon et al. (2017) and Myrvang et al. (2007) had raw material capacities greater than 500 tons per day. Generic data for plants of this size indicate energy requirements[1] of 1 520 MJ (with additional waste heat recovery) to process one ton of raw material (FAO, 1986), compared to 1 498 MJ and 1 890 MJ for the specific factories. Thus, there are differences between the actual energy requirements of existing fishmeal factories with similar throughputs, and the predicted typical value.

Fishmeal factories are encouraged to improve energy efficiency, due to the high energy requirements of the process and the negative environmental impact of fossil fuels. Fréon et al. (2017) reported that during fishmeal production, the use of fossil fuels was the single factor with the largest negative impact on the environment. Recommendations to reduce the amount of heat required were: to reuse steam within the factory, to eliminate steam leaks and to ensure optimal heat transfer by frequent descaling.

3.2 Background to the factories studied

Fishmeal factories are good candidates for ST heat integration in South Africa since they have a large demand for heat below 200°C and the more constant and predictable cost of solar energy could reduce some of the volatility experienced in this industry. Two fishmeal factories (Factory A [FA] and Factory B [FB]) along the western coast of South Africa were studied, due to the prevalence of the

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[1] Calculated from fuel consumption assuming a calorific value of 40 MJ/kg for heavy fuel oil.

fishing industry in the Western Cape and weather conditions in this area are favourable for ST heat integration. The Western Cape is responsible for most of the fishing activity in South Africa, accounting for an estimated 90% of the fishing industry (Hara et al., 2008). Solar energy was identified as one of the renewable energy sources with great potential in the Western Cape (DoE, 2015), due to the relatively high solar irradiation levels.

The two factories studied, were selected based on their distinct production regimes and throughputs. FA employs a unique production method and has a low throughput, it produces fishmeal mostly from by-products from hake and other white fish processing. Reductions in fishing quotas for South African demersal fisheries at the end of the twentieth century (Japp, 2001), resulted in a lack of raw material, thus, fishmeal production was downscaled and fishmeal sales were a minor source of income for FA. However, the increasing demand for fishmeal, along with the benefits of converting waste into a higher value product, have motivated FA to reconsider the importance of fishmeal production in its value-adding operations. Subsequently, fishmeal production has significantly increased due to renewed efforts to collect and process fish by-products and offal. With the greater production throughput in the factory, FA became aware of the various inefficiencies present in their facility, consequently, a strategic aim to decrease energy and scarce resources consumption arose.

FB has a much higher throughput relative to FA and produces both fishmeal and fish oil with the conventional wet-pressing method, using exclusively pelagic fish caught off the West Coast of South Africa as raw material. The large production amounts result in considerable energy requirements, which is currently satisfied by fossil fuels, the majority being coal. Fossil fuels constitute a large portion of the production costs of FB, therefore, reductions in fuel consumption could result in significant cost savings for the factory.

The use of fishmeal factories as case studies was beneficial both to this study on ST heat integration in the South African industry and to the specific factories studied. The two factories represent industrial energy users with different production methods, schedules and throughput rates, this enabled different types of factories to be investigated. Both factories identified a need for process optimization with regards the energy use, which lead to them participating in the study. Including renewable energies and improving the overall efficiency holds primarily two advantages for production facilities: 1) the direct and indirect financial burden of combusting fossil fuels can be reduced; 2) higher efficiency could result in improved equipment operation and less downtime, which is beneficial to production.

3.3 Base case process definition and characterisation methodology

Prior to investigating energy efficiency measures for an industrial process, an accurate model containing all the relevant process information is required in order to have a standard case against which any alterations to the process can be measured. This model, the base case process, should be representative of the actual process in its most recent configuration (Mateos-Espejel et al., 2011).

Overview of methodology

The methodology followed to define and characterise the base case processes is described in Figure 3-1. Site visits to the respective factories were performed to characterise the process and to collect production data, after which the data sets were processed to identify and remove inconsistent entries and obtain useful information. Based on observations during the visits, process flow diagrams (PFD's) of the processes were constructed, then the edited data sets were used to set up mass and energy balances. The PFD of each facility and the initial mass and energy balances were used to automate the calculations of mass flow and energy requirements in spreadsheets using Microsoft Excel 2016. These automated calculations represented simulations of generic processes with the same topology as those studied, these were updated with actual factory data and were then viewed as representative of fishmeal production at the respective facilities. With the energy requirements of the base case process known, the steam production system of each factory was simulated using Aspen Plus V8.8 process simulation software. Combined, these two simulations represented the base case process of each factory and enabled calculating the energy and fuel requirements. These simulations could then be used to quantify the effect of changing operating conditions within the process.

Collect data	• The factory is observed under operating conditions. • The process topology is established and equipment information collected. • Data on production amounts, operating conditions and fuel consumption over a period are collected.
Data processing	• Data is analysed and erroneous entries removed. • Average production amounts and typical operating conditions are determined from the data.
Construct PFD	• A process flow diagram is developed for the specific production process. • Additional sources are consulted for more information on equipment if necessary. • Mass and energy balances are performed manually.
Generic simulation	• Based on the PFD of the specific process, mass and energy balances are automated using spreadsheets.
Base case simulation	• The generic simulation is updated with average production information. • The results from the base case simulation are compared with known literature values and the actual factory data to validate the simulation.
Boiler simulation	• The steam production system is simulated with process simulation software to determine the steam flow rate required to deliver the energy required by the base case production process.

Figure 3-1: Methodology used to define and characterise the base case process of each factory studied.

The two factories studied differed significantly with respect to production schedule, production method and throughput, and the development of the base case processes presented different challenges. The ways in which these challenges were addressed are unique to each factory and are results in themselves, therefore, they are presented later in this chapter with the other results.

Processing collected production data

The collected production data were checked for errors and the useful information that could be extracted from it identified. Production data are captured per shift by both factories; however, after the data are captured no further validation or checking is performed. Each data entry consisted of multiple values, for example: the amounts of raw material and fuel consumed, the total production time, the amount of fishmeal produced and fishmeal properties like moisture content. In order to avoid the data set from becoming biased when removing individual values from a data entry, it was

decided to completely remove a data entry from the data set if one of its values was thought to be erroneous. The collected data sets were large enough and covered a sufficient period of time to enable this approach. Erroneous data entries were identified by applying the following criteria:

Data entries were deleted for the following reasons:

- If any value was negative, since the values recorded represent absolute amounts (e.g. mass of product or volume of fuel consumed)
- Where the production data were corrupted and the production time could not be calculated
- Where values were abnormally large, which was defined as greater than the mean + three times standard deviation
- Where a mass balance was not satisfied, i.e. mass being created in the process
- Where an entry was incomplete

Calculating the energy required during cooking and drying

To calculate the total energy requirements of the factories studied, the energy that needs to be transferred to the raw materials during cooking and drying was required. The energy transfer rate ($\dot{Q}_i$) required to heat a pure component stream (i) form an initial (T_1) to final (T_2) temperature, without a phase change occurring, can be calculated from the mass flow rate ($\dot{m}_i$) and specific heat ($C_p(T)_i$), by using Equation 3.1 (Ghajar and Cengel, 2014).

$$\dot{Q}_i = \dot{m}_i \int_{T_1}^{T_2} C_p(T)_i \, dT \qquad \textbf{Equation 3.1}$$

Choi and Okos (1986) developed polynomial models for the thermal properties of the most common food components: ash, carbohydrate, fat, fiber, protein and water. The generic form of the temperature dependent specific heat ($C_p(T)_i$) is shown in Equation 3.2. Coefficients to calculate the specific heat of the components present in fish materials (ash, fat, protein and water) are shown in Table 3.2.

$$C_p(T)_i = a_i + b_i T + c_i T^2 \qquad \textbf{Equation 3.2}$$

Table 3.2: Coefficients for the models of specific heat [kJ/(kg.K)] of food components, valid from -40°C to 150°C unless otherwise stated. Data from ASHRAE (2014).

Component	a	b	c
Ash	1.0926	1.8896×10^{-3}	-3.6817×10^{-6}
Fat	1.9842	1.4733×10^{-3}	-4.8008×10^{-6}
Protein	2.0082	1.2089×10^{-3}	-1.3129×10^{-6}
Water (0°C to 150°C)	4.1289	-9.0864×10^{-5}	5.4731×10^{-6}

The energy required to change the temperature of fish materials was calculated with Equation 3.3, using the models by Choi and Okos (1986). The exact value for specific heat was used by substituting Equation 3.2 into Equation 3.1 and performing the integration. It was assumed that the different components do not interact with one another and that the total heating energy required is the sum of the energies required by the individual components.

$$\dot{Q}_i = \dot{m}_i \left(a(T_2 - T_1) + \frac{b}{2}(T_2^2 - T_1^2) + \frac{c}{3}(T_2^3 - T_1^3) \right)$$ **Equation 3.3**

Equation 3.3 is sufficient for cooking processes since no evaporation occurs. In cases where evaporation occurs, such as drying, the heat of evaporation ($h_{fg,i}$) of the stream evaporated ($\dot{m}_i$) must be accounted for and additional energy will be required ($\dot{Q}_{evapo,i}$), see Equation 3.4. The heat of evaporation was obtained from the steam property tables provided by Çengel and Boles (2011).

$$\dot{Q}_{evap,i} = \dot{m}_i h_{gf,i}$$ **Equation 3.4**

3.4 Factory A base case process definition and characterisation

The production process and resource requirements of FA is described in this section, as well as the base case process definition and characterisation. The results of FA's base case process are summarised and compared to that of FB and literature values in section 3.6.

3.4.1 Description of data collected from Factory A

A preliminary visit to FA on 11 February 2016 was used to meet key personnel and get an overview of the facility. Then on 28 and 29 September 2016 data collection, interviews with technical personnel and observation of the factory during production were performed.

FA records data at the end of production runs in the form of weekly reports, containing critical parameters like the amounts of raw material and fuel consumed, fishmeal produced and average product moisture. A production run refers to a continuous period of fishmeal production, lasting until the stored raw material has been depleted. Weekly reports for runs ending on 19 September 2015 to 24 September 2016 were included in this study. Of a total 54 production runs achievable during this period, 47 complete reports were available, which accounts for 87% of the possible runs. It was decided that the collected data were sufficient to accurately characterise the performance of the production process at FA. The production data collected for FA was over a sufficient period (of one year) to characterise the long-term behaviour of the process (Mateos-Espejel et al., 2011). The weekly report data collected from FA are shown in Appendix B.

When FA is in operation, operating conditions are recorded at roughly thirty-minute intervals on process control sheets. Process control sheets for operation between 20 September 2016 and 28 September 2016 were provided by the facility to be used in this study, which represents approximately 133 total hours of operation and contained 219 data entries. FA operated normally during this period, according to reports by factory personnel. Process control sheet data collected from FA are shown in Appendix B.

In addition to the weekly production reports and process control sheets, a compositional analysis (provided by the factory) of the fishmeal product in September 2016, and various in-house technical reports and manuals were used to define and characterise the base case process of FA.

3.4.2 Factory A production process and ancillary systems

The production mode at FA is semi-batch, with a single continuous production run of approximately 5 days occurring roughly on a weekly basis, depending on the amount of raw material stored over the preceding week. Production is initiated once sufficient amounts of raw material have been collected, and once initiated, FA is run continuously for 24 hours per day until all raw materials have been depleted, at which time production is terminated and the factory shuts down.

Fishmeal production process

A process flow diagram of the fishmeal production process of FA was constructed and is shown in Figure 3-2; together with the description in this section, this defines the production process at FA during the period it was studied. FA transforms raw fish material into fishmeal as follows:

   i. The raw material, lean fish trimmings and offal, are collected in a temperature controlled storage room and kept between 8°C and 12°C before being processed.

   ii. Raw material is loaded onto a speed-controlled conveyer which transports it through a metal detector, ejecting any metal containing raw materials.

   iii. Metal free raw material is transported by a sealed screw conveyer to a wet mill which ensures particle sizes of 10 mm or smaller.

   iv. Screw conveyers transport the milled material to the dryer, two of the conveyers are fitted with stainless-steel sleeves that enable the material to be heated with steam condensate exiting the dryer.

   v. The dryer (TST-80R manufactured by Atlas Stord) is a single unit that simultaneously cooks and dries the materials. The internal construction consists of discs which are heated to temperatures in excess of 110°C, using steam at 500 kPa (gauge). Vapours from the dryer are extracted and condensed in a condenser using sea water.

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vi. The dried material exits the cooker and is transported to a hammer mill with a screw conveyer, sea water in a stainless-steel sleeve cools the fishmeal.

vii. The hammer mill reduces average particle size and disrupts agglomerated material formed during drying. The mill is connected to an air extraction system, a bag filter is used to capture dust resulting from milling before the air enters the condenser. Milling is the final step in the production process, the result being fishmeal in fine powder form.

viii. A chemical scrubber, that uses chlorine dioxide, treats the effluent from the condenser before it is expelled from the factory, to reduce unwanted odours.

ix. A screw conveyer transports the fishmeal to a screen where oversized particles and foreign material are removed before packaging in 25 kg or 50 kg laminated hessian bags.

x. The packed fishmeal is stored until it is shipped, which takes place roughly twice a month.

The raw material utilised by FA has a very low oil content, therefore, no significant amounts of oil need to be separated during the process, hence no oil separation equipment form part of this process. The production method used by FA is similar to the production without cooking method, described by FAO (1986), with the further simplification that only a single dryer is employed. In this case it is an acceptable process modification, since the throughput of FA is relatively low and the factory processes only lean (low oil content) fish, and this method is commonly used in such factories.

FA does not supplement any preservatives or antioxidants to the raw material since the raw materials are maintained at low temperatures from point of capture until the raw materials are fed to the fishmeal plant, which ensures high quality materials. FA does not use antioxidants in their fishmeal, due to small amounts of oil present in the final product.

Ancillary systems

The steam production system of FA is included in Figure 3-2. The boiler at FA was manufactured by John Thompson, and it has a steam production capacity of 5 metric tons/h. Heat is supplied to the boiler by combusting a heavy fuel oil (HFO) blend. An electrical heater with a recirculation circuit is used to heat the HFO to 60°C, thereby reducing the viscosity and making flow easier.

The boiler produces steam at 800 kPa (gauge) and has a maximum pressure rating of 1 000 kPa (gauge). The steam is throttled to 500 kPa (gauge) with a valve prior to entering the dryer. A steam trap ensures only condensate exits the dryer, condensate exits at 150 kPa (gauge). A fraction of the condensate is used for preheating the materials entering the dryer while the rest is directly sent to the hot well tank. Fresh water replaces condensate lost by leaks. Water exiting the hot well tank is treated with chemicals and pumped to the boiler with a multistage centrifugal pump (Grundfos, model A96501221).
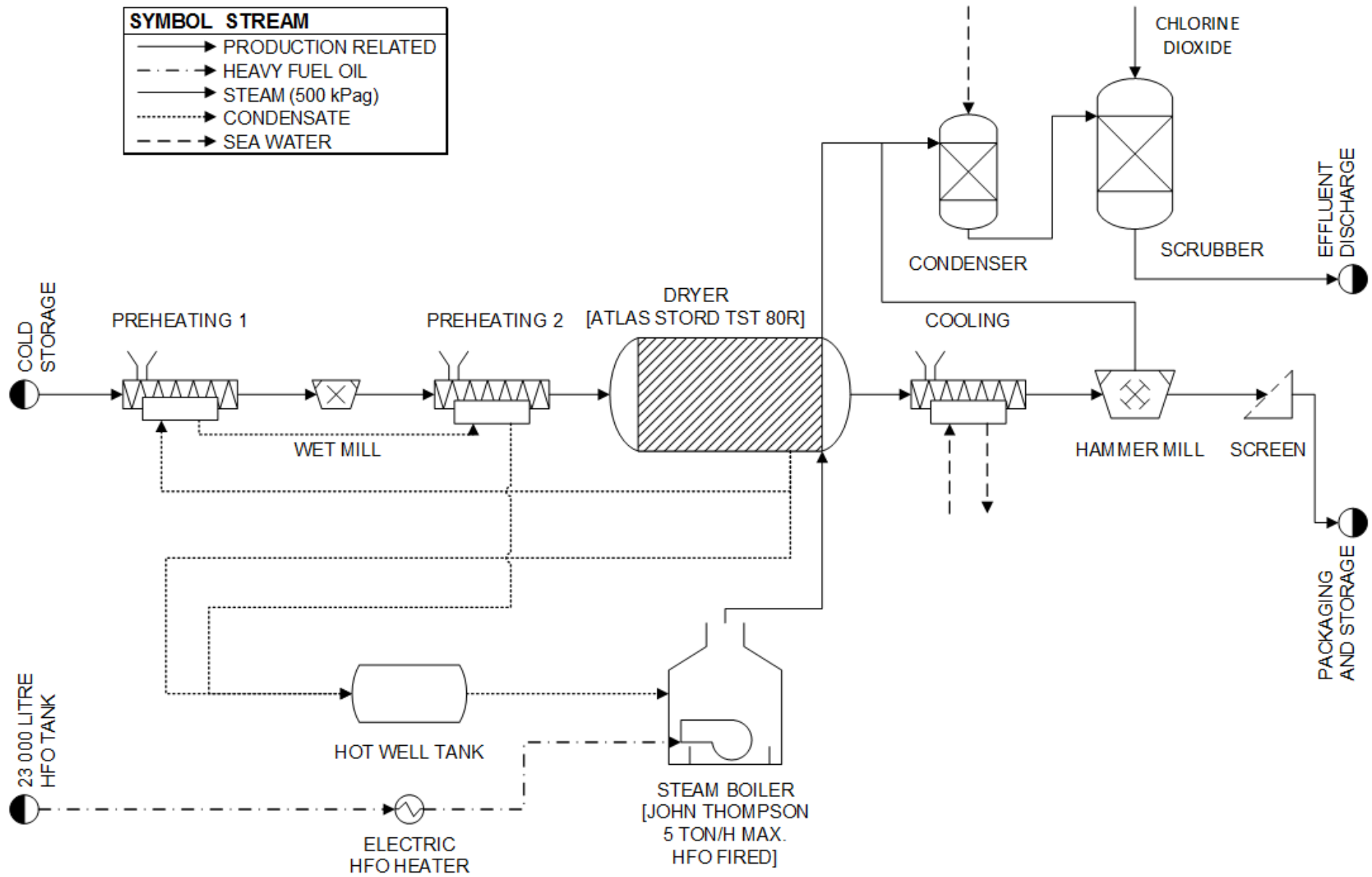
Figure 3-2: Process flow diagram of the fishmeal and steam production processes of Factory A.

3.4.3 Factory A typical production rate

With the production process defined, the typical production rate was determined. The raw material mass flow rate was selected as the independent variable, with the fishmeal product mass flow rate calculated from the yield.

Total annual production at Factory A

FA has a low annual production capacity and a yield that is typical for the fishmeal industry in South Africa. From September 2015 to September 2016, FA produced 1 441 metric tons of fishmeal from 6 235 metric tons of raw material, which is very low compared to specific factories in Norway (Myrvang et al., 2007) and Peru (Fréon et al., 2017) with an annual raw material capacity in excess of 100 000 metric tons. The average yield of fishmeal from raw material was $23 \pm 2\%$, which corresponds well with the reported average yield of 23% for South African factories (de Koning, 2005).

Weekly production at Factory A

The instantaneous raw material feed rate needed to be calculated in order to characterise the base case process. Since the weekly production reports cover a period limited to the duration of a single production run, they were used as a starting point to determine the throughput rates of the factory. A typical production run was characterised by averaging the values in all available weekly production reports, and the values are shown in Table 3.3. The weekly production target set by FA was 35 metric tons, further, the actual production was 31 ± 9 metric tons, with a minimum of 15 metric tons and a maximum of 62 metric tons. On average, the actual weekly production was less than the target set by FA, however, the relatively large variation means that production was frequently greater than the target.

The variation in the weekly production is believed to be due to fluctuating raw material availability. The standard deviation of the amount of raw material consumed was approximately 30% of its mean value, this was also the case for fishmeal produced (29%) and fuel consumed (34%). These values indicate that the variation is due to the varying amounts of raw materials consumed, this is true for the fishmeal produced since the yield showed lower variance than the other parameters, with the standard deviation approximately 9% of its mean.

Table 3.3: Average values for Factory A production runs, calculated with values for 47 runs, over the period September 2015 to September 2016.

Parameter	Value (mean ± standard deviation)	Unit
Fishmeal produced	31 ± 9	Metric ton
Raw material consumed	133 ± 40	Metric ton
Yield	23 ± 2	%
Fuel consumed	10 018 ± 3 362	Litre

Variation throughout the year

The yield obtained for each production run remained fairly constant, despite the amount of raw material processed varying between production runs. Figure 3-3 shows the yield obtained and raw material consumed per production run at FA from September 2015 to September 2016. The yield remained relatively constant, which indicates that the raw material quality and composition remained constant throughout the period. There does not appear to be any seasonal variations in yield or amount of raw material consumed, thus, variation was assumed to be random, caused by the fluctuating availability of raw materials. Furthermore, certain individual production runs were longer than the rest and therefore utilised more raw material during those specific runs.
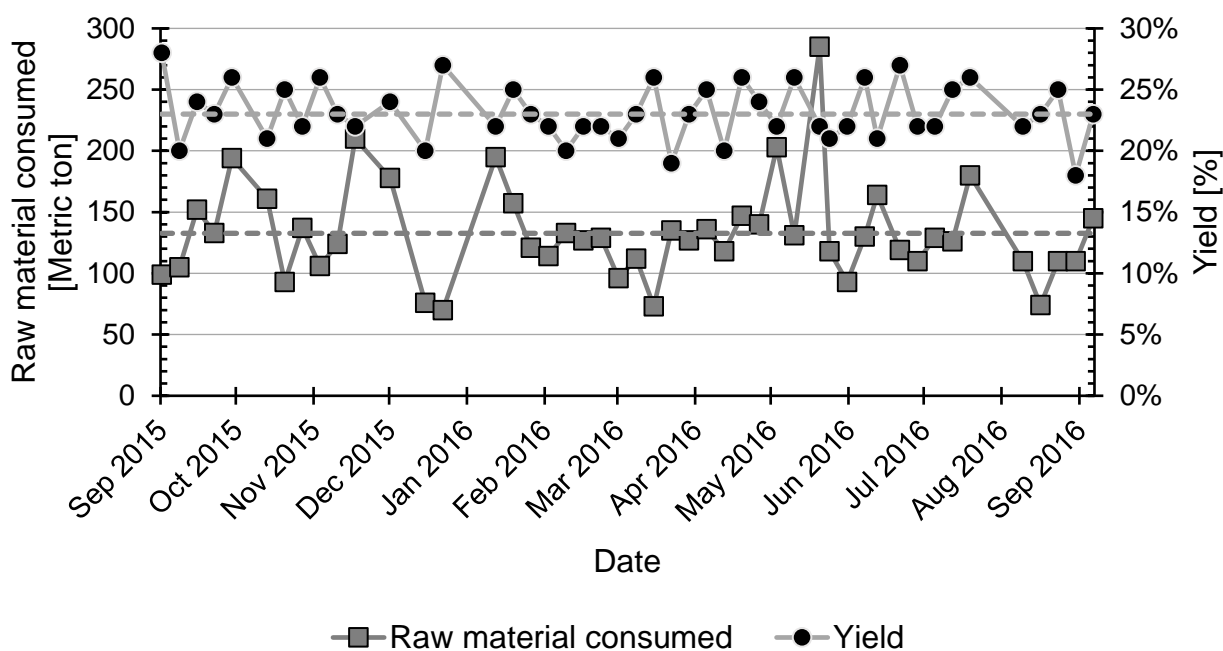
Figure 3-3: The mass of raw materials consumed and yield obtained for each production run at Factory A from 19 September 2015 to 24 September 2016. Average values are indicated with horizontal lines.

Although all fishmeal factories experience fluctuations in the amount of raw materials processed, FA can mitigate this to some extent by purchasing raw materials from various sources. Since fluctuations in production amounts are random, it is believed that, if the raw materials were available, production would remain constant throughout the year. Therefore, raw material consumption per production run for the base case process was defined as 133 metric tons.

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Duration of a production run

With the mass of raw material consumed per production defined, the duration of a typical production run was required to calculate the mass flow rate of raw materials to the fishmeal production process. According to factory personnel production at FA occurs throughout the year, with no scheduled holiday shutdown periods, and the factory is licensed to operate for 260 days per year for 24 hours per day. Based on this schedule, there are 6240 hours of production per year, for 52 weeks per year, this results in 120 hours per production run. However, it is known that time is lost due to start-up and shut-down, and the factory is not operating at maximum capacity, thus, a typical production run only produces product for approximately 100 hours. This is similar to durations recorded on the process control sheets. Therefore, the duration of a production run was defined as 100 hours of continuous operation for the base case process.

Raw material and product mass flow rates

The average mass of raw material consumed per run, 133 metric tons, and the production run duration of 100 hours were used to calculate the mean raw material mass flow rate: 1 327 kg/h. The mass flow rate of fishmeal produced was calculated as 307 kg/h by using the average yield of 23.2%. The only separation in this process is by evaporating water in the dryer, thus, 1 019[1] kg/h of water is removed.

3.4.4 Composition of Factory A raw material utilised and product

Composition of fishmeal product

FA frequently has the fishmeal product sent for compositional analysis by an independent laboratory; the ash, fat, water and protein content were included in the base case process. The compositional analysis for fishmeal produced at the start of September 2016 was used, this is shown as part of the mass balance in Table 3.4.

FA produces a high-quality fishmeal, which is classified as standard quality due to the high ash and slightly lower protein content than that specified for superior grade. According to the classification used by FA, this fishmeal would be standard grade since the protein content of 66.8% is slightly less than the 67% required for superior grade. The ash content is similar to the 18% of standard grade, and the fat content at 10.1% is only slightly more than the 10% or less required for superior grade.

Raw material composition

The composition of all streams entering and exiting the dryer was required for energy balance calculations. The composition of the entering stream and the vapour exiting the dryer was unknown, as these are not directly measured.

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[1] Calculated using unrounded values of 1326.66 kg/h for raw material and 307.23 kg/h for fishmeal.

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One method of obtaining the compositions would be to sample the relevant streams and have the samples analysed. Variation in the raw material consumed by FA over time means that a single sample may not result in representative results, as it would only give the composition of the raw material at a particular point in time. To obtain a reliable long-term average of raw material composition through sampling would therefore require extensive sampling and incur significant analytical costs. For this reason, it is not standard practice to sample the raw material stream.

The composition of fishmeal produced remains relatively constant, despite changing raw material compositions. This is due to the process being manipulated to ensure a reasonably constant product moisture and the raw materials from different sources being mixed during production. Therefore, the product composition was used in a mass balance and the raw material composition back-calculated (see Table 3.4) from the available data, as this would result in a more representative long-term average raw material composition.

In order to back-calculate the raw material composition from the available fishmeal product composition it was assumed that the vapour stream exiting the dryer is water only. This assumption was reasonable since water is the major component present in the system and amounts of other components evaporated would be negligible compared to water. With the compositions of the fishmeal product and evaporate streams known, the composition of the incoming raw material stream was calculated with a mass balance (the mass flow rates are shown in Table 3.4).

The composition calculated for the raw material stream of FA (see Table 3.4) compared well with literature values for similar raw materials, therefore, it is believed that the calculated values are representative of the actual raw material utilised by FA. Hake was selected as a reference species to compare the calculated composition with, since FA mostly consumes by-products from hake processing. The average composition values for whole hake is: 16.8% protein, 2.0% fat, 3.0% ash and 78.2% water (FAO, 1986). The calculated fat and water content, at 2.3% and 78.0% respectively, compared well with the literature values for whole hake, while the protein was lower at 15.5% and the ash higher at 4.2%. The differences in protein and ash were expected since the offal and frames contain less flesh than whole hake, while most minerals are in the frames, which would increase the ash content.

Table 3.4: The composition of fishmeal produced at Factory A in September 2016, calculated composition of the raw material stream, and mass flow rates of streams entering and exiting the dryer.

Component	Raw materials		Vapour stream[1]	Fishmeal product	
	Composition [kg/100 kg]	Mass flow rate [kg/h]	Mass flow rate [kg/h]	Composition [kg/100 kg]	Mass flow rate [kg/h]
Ash	4.2	55.6	0.0	18.1	55.6
Fat	2.3	31.0	0.0	10.1	31.0
water	78.0	1034.8	1019.4	5.0	15.4
Protein	15.5	205.2	0.0	66.8	205.2
Total	**100.0**	**1326.7**	**1019.4**	**100.0**	**307.2**

3.4.5 Factory A typical operating temperatures

The temperatures that characterise the base case process were defined either from ranges reported by FA personnel, or the process control sheets. According to production guidelines the raw materials are stored between 8°C and 12°C. The lowest temperature of 8°C was used for subsequent calculations based on observations at the factory. Preheating the raw material stream with steam condensate exiting the dryer raised its temperature to between 40°C and 50°C according to factory personnel, who periodically take measurements of this stream. A mean temperature of 45°C was used in the base case process. The temperature of materials exiting the cooker is measured by a probe located at the outlet and the values recorded on the process control sheets. The average outlet temperature of 116.9°C was used in the base case, with the assumption that the fishmeal and water vapour exiting the dryer were at the same temperature.

3.4.6 Factory A base case process simulation

Fishmeal production process simulation

The fishmeal production process of FA was simplified to include only units that change the temperature or composition of material streams, that were relevant to the SHIP study and that sufficient information was available for. The simplified process consists of the preheating and drying stages and is illustrated in Figure 3-4. The cooling, milling and effluent treatment stages were not included as they do not consume energy from steam and were not relevant to the SHIP study presented in Chapter Four. The mass and energy balance calculations of this process were automated using spreadsheets in Microsoft Excel 2016, these were used to simulate the mass flows and energy

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[1] The composition of the vapour stream is not given since it was assumed that the evaporate stream is water only.

requirements. Images of the simulations of the preheating and drying sections are shown in Appendix B.
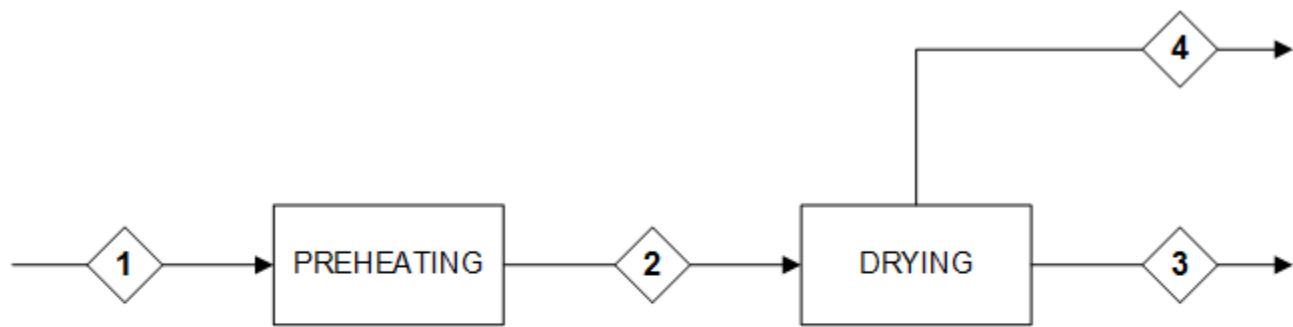
Stream Number	1	2	3	4
Temperature (°C)	8	45	117	117
Mass Flow (kg/h)	1327	1327	307	1019
Composition				
(Component Mass/Total Mass)				
Ash	4.2%	4.2%	18.1%	0.0%
Fat	2.3%	2.3%	10.1%	0.0%
Protein	15.5%	15.5%	66.8%	0.0%
Water	78.0%	78.0%	5.0%	100.0%

Figure 3-4: Flow diagram of the base case process simulation of Factory A. The process was simplified to include only units that affect temperature or composition changes.

Steam production system simulation

The steam production system including steam users: the dryer and raw material preheating circuit, of FA were simulated with Aspen Plus V8.8 process simulation software. These needed to be characterised for the base case process since FA does not measure the steam flow through the dryer, or the fraction of steam used for preheating. The flowsheet of the simulation in Aspen Plus V8.8 is shown in Figure 3-5, a description of the simulation follows:

i.   The steam circuit was simulated as a system with total recirculation, due to the low amount of steam losses at FA and the make-up water not being measured. Systems that include recirculation are difficult to simulate from an initialised sate (attempting simulation typically results in errors), therefore, a selector block (SELECT) and a separate input stream (5) were employed to enable simulating once-through flow for the system, instead of recirculation. The simulation was run with stream 5 selected, after each simulation the mass flow rate and temperature of stream 5 was updated with that of the exit stream (12) until the values converged, at which point recirculation within the system was simulated by selecting stream 12.

ii. The boiler produced saturated steam at 800 kPa (gauge); a design specification manipulated the steam mass flow to obtain the required net duty for the boiler.

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iii.   Steam was throttled to 500 kPa (gauge) with a valve (VALVE1).

iv.   Steam exited the dryer as condensate at 150 kPa (gauge). The energy change of the steam was assumed to be the energy transferred to the dryer.

v.   A fraction of the condensate was diverted (SPLIT1) and used for preheating of the raw material, while the remainder flowed directly to the hot well tank. Preheating of the raw material was simulated with a heater (PREHEAT) that removed the amount of energy transferred to the raw material to increase its temperature from 8°C to 45°C. A design specification manipulated the fraction of condensate diverted to preheating, to ensure a condensate temperature of 67°C after preheating. The outside temperature of the sleeves used for preheating was measured as 65°C and the temperature of condensate on the inside was assumed to be 2°C higher, since the sleeves were thin.

vi.   All pipes were insulated and had relatively short lengths; therefore, all heat losses to the environment were assumed to occur at the hot well tank since it had a large surface area and was not insulated. Condensate exited the HWT at 90°C, the heat losses to the environment were taken as that amount of energy that needed to be removed to result in a final temperature of 90°C.

vii.   The pump (Grundfos A96501221) increased the pressure by 1 312 kPa at the specific flowrate, obtained from the pump curve. The pump curve is shown in Appendix B.

viii.   The pressure of the stream exiting the pump (11) was throttled with a valve (VALVE2) to 800 kPa (gauge) , the operating pressure of the boiler.
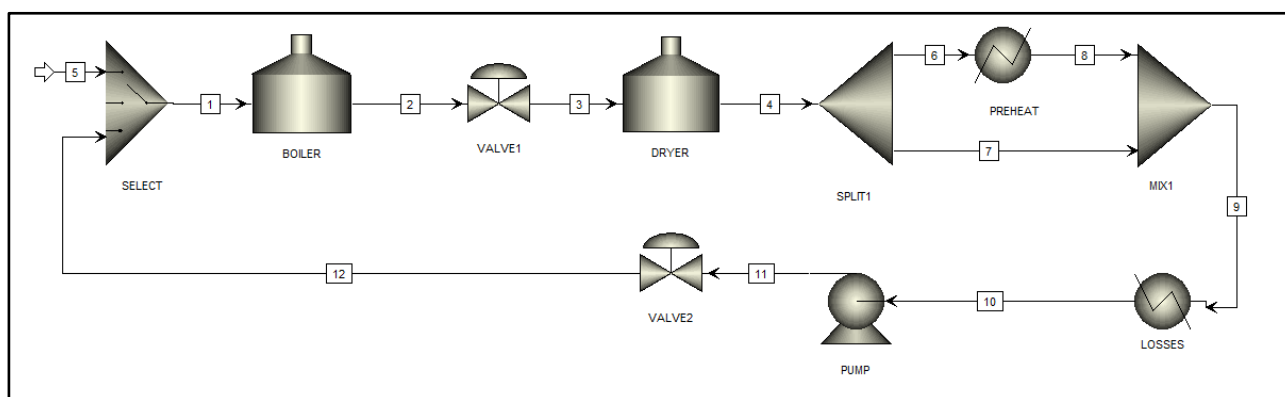
Figure 3-5: Aspen Plus V8.8 simulation flowsheet of the steam production system and steam users at Factory A.

3.4.7 Factory A base case process energy and fuel requirements

Energy required for preheating and drying

The drying step at FA was energy intensive, mostly due to the large amounts of water that had to be evaporated. With a raw material mass flow rate of 1 327 kg/h, drying the material required 2 659.2 MJ/h and preheating required 178.2 MJ/h. Preheating required significantly less energy than drying

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since no phase changes occurred during preheating. The total energy required and the energy required by the individual components are shown in Figure 3-6, water accounted for the greatest amount of required energy since it is the major component present and the only component to change phase during drying.
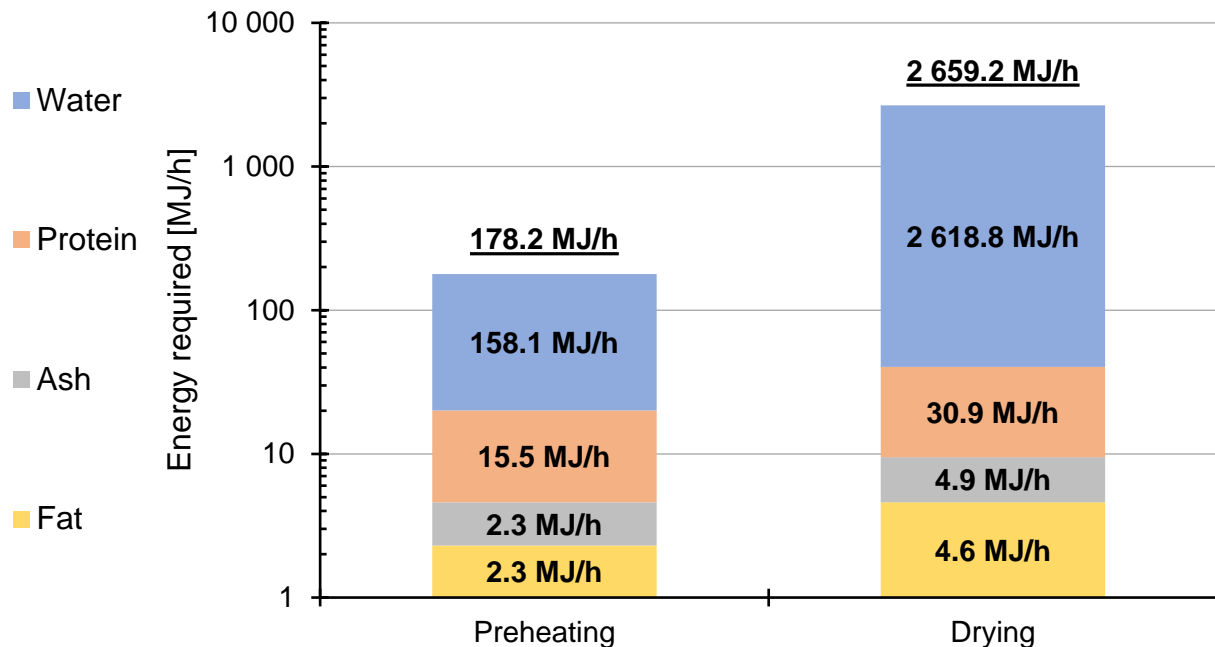
Figure 3-6: Energy required to preheat and dry 1 327 kg/h of raw material in Factory A. Preheating to 45°C and, drying to 117°C and 5% humidity occurred.

Calculating the required mass flow rate of steam to the dryer

To calculate the required steam mass flow rate to the dryer, the heat transfer efficiency as well as the energy transferred to the materials were needed. The heat transfer efficiency of the dryer was not known and is difficult to calculate accurately or obtain from literature since it depends on various factors such as equipment design, age, operating regime and frequency of cleaning and maintenance.

A more realistic method was to use the information available to calculate the efficiency of equipment for the period studied. The inlet and outlet conditions of steam in the dryer were known, thus the energy lost by the steam and correspondingly gained by the dryer unit was calculated.

A flow device between the hot well tank and the boiler was used to record the total volume of water to the boiler per production run. The average volume of water used per run was 214 ± 60 m^3, for a 100-hour production run the average volumetric flow rate was 2.14 m^3/h. The conditions of the stream at the flow meter were known, therefore, the simulation of the steam system in Aspen Plus V8.8 was adapted to replicate the conditions at this point and obtain missing information, such as the heat losses to the environment and the fraction of steam used for preheating the raw material.

Steam production based on volumetric flow rate of water

From the simulation of the steam production system in Aspen Plus V8.8, the required mass flow rate of steam and energy required by the boiler was obtained and consequently the fuel consumption was calculated. In the simulation of the steam system, the mass flow rate of water was manipulated to obtain a volumetric flow rate of 2.14 m^3/h at the outlet of the hot well tank, with important results from the simulation shown in Table 3.5. Producing the 4 865 MJ/h of energy for the boiler, required combusting 136.6 litre/h of HFO. A net calorific value of 39.57 MJ/litre (Global combustion systems, 2017) for HFO was used, with a boiler-burner efficiency of 90% defined; the efficiency provided for similar oil fired boilers manufactured by John Thompson (John Thompson, 2017). This was a reasonable assumption since the boiler at FA is inspected every three years and the burner is serviced at monthly intervals.

Table 3.5: Results of the steam production system simulation of Factory A in Aspen Plus V8.8, based on a volumetric flow rate of 2.14 m^3/h to the boiler feed pump from the hot well tank.

Parameter	Value	Unit
Mass flow rate of steam	1 991	kg/h
Proportion of condensate redirected for preheating	32.46	%
Heat required by boiler to produce steam	4 865	MJ/h
Required HFO flow rate	136.6	litre/h

The required HFO flow rate of 136.6 litre/h does not correspond to the flowrate calculated from the actual data collected onsite: calculations from collected data give a value of 100.2 litre/h HFO per 100-hour production run (see Table 3.3). Combustion at this rate releases 3 964 MJ/h, thus, even if the boiler was operating at 100% efficiency, this fuel flow rate would not be sufficient to provide the energy required to produce 1 991 kg/h of steam.

The HFO flow rate calculated from the simulation, based on the recorded flow rate of water, was 36% higher than the actual fuel flow rate calculated from FA data. Since FA is invoiced, and pays, per litre of fuel, it is highly unlikely that the recorded HFO consumption was incorrect. The water in the steam production circuit continually recirculates, and if minor losses due to leaks are replenished, the exact flow rate does not affect the overall process. Therefore, it is most likely that the flow measurement device at the HWT was faulty and the recorded water use was incorrect. Furthermore, there are significant differences between the specific water usage per ton of product reported by FA and those calculated by the author. Therefore, it was assumed that the water usage data obtained from the flow meter were unreliable and the base case process steam use was characterised based on the average fuel flow rate calculated from FA data.

Steam production based on volumetric flow rate of fuel

The actual HFO flow rate of 100.2 litre/h was used, and the simulation of the steam production system in Aspen Plus V8.8 adapted to ensure 3 568 MJ/h of energy transferred to the steam generated in the boiler. This is the energy transferred to the boiler by HFO combustion at 100.2 litre/h for a boiler-burner efficiency of 90%. The mass flow rate of steam was manipulated to obtain this heating duty for the boiler.

The steam flow for the base case process was defined as 1 460 kg/h, which required combusting 100.2 litre/h of HFO, and resulted in 44% of the steam condensate being redirected to preheat the incoming raw material stream. The results for the simulation based on the actual average fuel flow rate is shown in Table 3.6. This scenario based on the average fuel consumption was assumed to be most representative of the current mass and energy flows at FA, due to the following reasons:

i.   This scenario results in the same average fuel use as in reality, which is one of the major factors affecting the cost of production.

ii.   A steam mass flow rate of 1 460 kg/h corresponds to 4.75 kg steam being used for each kg of fishmeal produced, which is within the range stated by FA personnel.

Table 3.6: Results of the steam production system simulation of Factory A in Aspen Plus V8.8, based on a net heating duty of 3 568 MJ/h for the boiler, representing 100.2 L/h of HFO being combusted with a 90% energy efficiency.

Parameter	Value	Unit
Mass flow rate of steam	1 460	kg/h
Proportion of condensate redirected for preheating	44.26	%
Heat required by boiler to produce steam	3 568	MJ/h
Required HFO flow rate	100.2	litre/h

Equipment efficiencies

The heat transfer efficiency of the dryer at FA was calculated as 80%, which is relatively high, however, the value seemed reasonable since the dryer was designed for fishmeal production. Steam enters the dryer at 500 kPa (gauge) and exits as condensate at 150 kPa (gauge), and for a steam mass flow rate of 1 460 kg/h (the base case process flow rate), 3 320 MJ/h of energy was transferred to the dryer. With 2 659 MJ/h of this energy effectively transferred to the fish material, the quotient of these values gives the heat transfer efficiency of the dryer as 80%. The heat transfer efficiency of the sleeves used to preheat the materials was assumed to be 100%, due to the small amount of energy transferred compared to drying and a lack of information required to calculate the efficiency.

3.4.8 Factory A yearly production schedule

<u>Characterising the base case schedule</u>

In addition to the hourly production rates, the yearly production schedule was characterised and a schedule defined for the base case process. FA stated that production typically starts on Tuesdays; from information on the process control sheets it was assumed that production starts at 11:00 am every Tuesday, the 100-hour production run continues until 03:00 pm on Saturday. This weekly production schedule was applied to the entire year of 2016, since there are no shutdown periods for fishmeal production at FA, and thus the yearly production schedule for the base case process was defined.

<u>Comparing the base case process schedule with reality</u>

An annual production schedule was set up for FA, the purpose of the schedule was to characterise the long-term average operation of the factory and eliminate once-off occurrences, while still being representative of the factory in terms of the available data. The annual production schedule was described in terms of the amount of raw material consumed per month, the calculated values for the base case process and the actual amounts consumed during 2016 are shown in Figure 3-7. As expected, the base case amounts show less variation as these were calculated from average values. Actual production is dependent on raw material availability, which in turn is dependent on fishing schedules and weather patterns. However, even though differences between the base case process and actual production were significant in months like January, June and August, there appears to be a good comparison between the base case schedule and the actual production at FA for the majority of the production year.
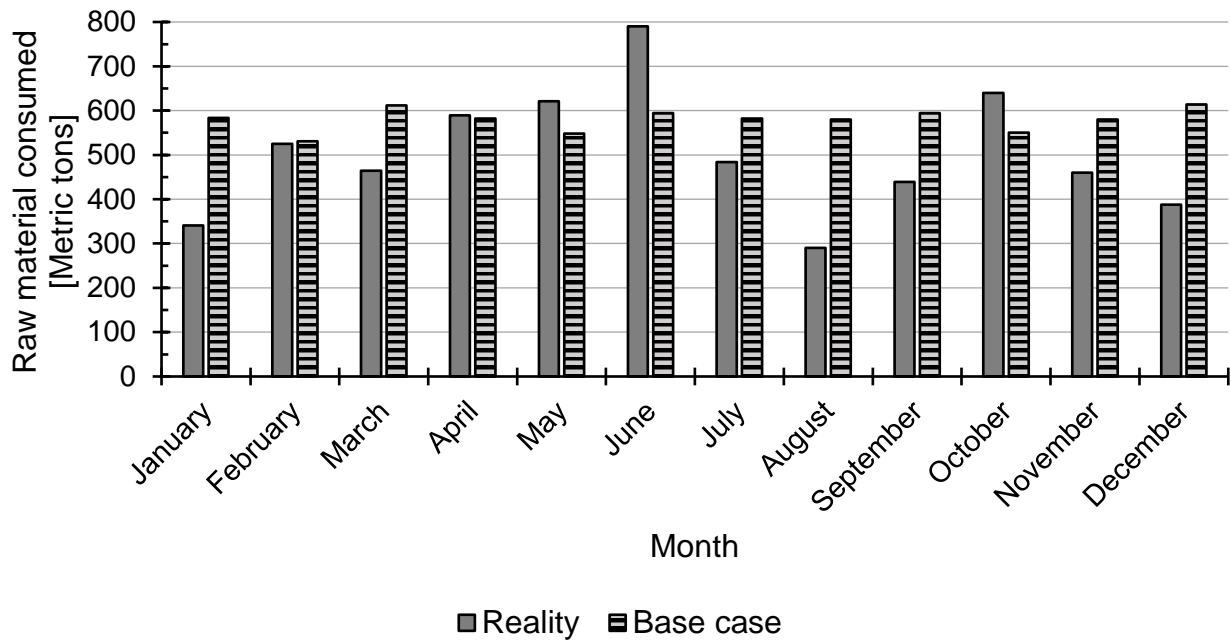
Figure 3-7: Raw material consumed per month for Factory A during 2016, both the actual and base case process values are shown.

3.5 Factory B base case process definition and characterisation

The production process and resource requirements of FB is described in this section, as well as the base case process definition and characterisation.   The results of FB's base case process are summarised and compared to that of FA and literature values in section 3.6.

3.5.1 Description of data collected from Factory B

A site visit to FB was completed on 2 to 5 May 2017, during which the factory was observed while producing FMFO, production data were collected and factory personnel were interviewed. Data for the 2016 production period, from 15 January to 16 December 2016, were obtained. Production data for FB were available in the form of worksheets created after each shift and daily production summaries. Steam production data for the boiler house were available in the form of worksheets completed after shifts. Compositional analyses for fishmeal and fish oil produced during January 2017 were available. The daily summary data and steam production data are shown in Appendix C.

Data on operating conditions from 13 January 2016 to 22 June 2017 were obtained, data were recorded in minute intervals and stored in a separate file for each day. The files were stored in a proprietary format and a program provided by FB was used to access the data per file. These data contained all the variables monitored by the programmable logic controllers (PLC's) used to control the production system equipment, and the most useful data were the outlet temperatures of the cookers and dryers.

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The fishmeal and steam production data collected from FB were processed and erroneous data entries, which had missing or corrupted (spreadsheet entries unreadable by Microsoft Excel 2016) values, or values which caused a mass balance to be violated, were removed. Sufficient data entries remained to be representative of the factory, further the errors for shift data were scattered throughout the year. The number of valid entries that remained and the percentage of data entries that were discarded are shown in Table 3.7. The shift data for fishmeal production had the most errors and almost half of the data were removed, while most daily and boiler production data remained.

Table 3.7: Summary of data sets collected from Factory B.

Data set	Original number of entries	Remaining valid entries	Percentage discarded
Shift data	441	239	45.8%
Daily data	89	75	15.7%
Boiler data	201	179	10.9%

The shift and daily production data sets represented the same production run, the difference between the two being the period over which the data were reported. Therefore, the sum of shift data for a specific day should equal the daily data for that day, however, it was not possible to reconcile these two data sets, neither before nor after removing erroneous data. However, the individual data sets still contained data that provided valuable insights into the operations at FB. The shift data set contained raw-material-specific information and the daily data contained information on the overall throughput rates and yields of FMFO.

3.5.2 Factory B production process and ancillary systems

Fishmeal and fish oil production process

Based on observations during the site visit and in-house technical documents, PFD's of the production process at FB were created. In accordance with the different sections defined in FB, Figure 3-8 shows the PFD of the 'wet' section (characterised by separation of liquids and solids by mechanical means, and streams with relatively high water content) and Figure 3-9 shows the PFD of the 'dry' section (characterised by removal of water by evaporation, and streams with low water content). At unit operations where separation occurred, the exit stream with the relatively higher solids content was labelled as 'mostly solids' and the other stream as 'mostly liquids', even if the major component present in both streams was water. The boilers, steam and condensate lines, and the vapour extraction and treatment equipment are not shown on the PFD's, in order to simplify the flowsheets and clearly show the major material flows. The discussion in this section, combined with the PFD's, describe production at FB during the time it was studied.

The fishmeal and fish oil production process at FB is as follows:

i. Pelagic fish are received at the harbour of FB, a pneumatic suction pump transfers the fish from the vessels onto a conveyer belt which transports it to storage at the factory. The raw material reaches ambient temperature before it is processed since the storage facilities are not actively cooled.

ii. Screw conveyers transport the raw material from storage to a hopper that feeds the cookers, formalin is added as a preservative during this transport stage.

iii. The raw fish pass through a magnetic separator to remove any metal objects.

iv. FB contains three non-identical cookers, and any combination of the three cookers may be used at any one time during production. Two of the cookers are followed by strainers and presses, while the other one is connected to a tank that feeds the decanters. The cookers use steam at gauge pressures of 120 kPa (123°C) to 300 kPa (144°C).

v. Cooked materials are strained and pressed by twin-screw presses to remove liquids liberated during cooking, liquids are sent to a tank and the press cake is transported to the dryers by screw conveyers.

vi. Horizontal decanter centrifuges are used to remove suspended solids from the press liquor, or the cooked materials from cooker 1, when in operation. The decanters produce a stream that is relatively higher in solids, known as a grax, and a liquid stream. The grax joins the press cake traveling to the dryers, while the liquid stream is sent to centrifuges.

vii. The liquid stream from the decanters is centrifuged to separate the oil and stickwater. The oil is polished in a small centrifuge and pumped to storage tanks, and the stickwater is concentrated.

viii. The water evaporated in the dryers drying the solids is used to concentrate stickwater in multiple-effect evaporators. The stickwater concentrate is sent to a tank before it is mixed with the press cake and decanter grax, and dried.

ix. Drying is performed in two stages consisting of three dryers in parallel for each stage. The press cake, decanter grax and 60% of the stickwater concentrate are dried in the first drying stage. The product of the first drying stage and the remainder of the stickwater concentrate are dried in the second stage. The dryers use steam at 550 kPa (gauge) (162°C).

x. A final check with magnetic separators, to ensure no foreign metal objects are present, is performed before the dry fishmeal is milled, with the milling stage consisting of six hammer mills in parallel.

xi. After milling, ethoxyquin is added to the fishmeal to prevent oxidation of the oils present. Fishmeal is transported to the packaging section, and packaged in either 20 kg or one-ton bags before being despatched.

xii. A ducting system is installed throughout the entire factory to remove vapours and steam from most equipment pieces and conveyers. The vapours are sent to a scrubber and treated before being discharged into the sea.

FMFO production at FB is an example of the wet-pressing method that is commonly used internationally. Autolysis of the raw material in the receiving storage facility, exacerbated by relatively long waiting times and high temperatures, requires the use of formalin as a preservative to ensure appropriate material properties in the process. The relatively high oil content of the fishmeal, at 11 kg/100 kg fishmeal, requires the use of ethoxyquin as an antioxidant to avoid spoilage of the fishmeal by oxidation of the oils.

FB uses two kinds of dryers: Atlas Stord TST 90 SP dryers, which are typical disc dryers with steam flowing through a hollow shaft with discs heating the material, and taç makina D10 000 dryers which are rotary drum dryers with steam flowing through the outer shell heating the material in contact with the inside walls.

Ancillary systems

The steam required in the fishmeal production process is obtained from a central steam production system servicing multiple facilities. Steam is produced at 800 kPa (gauge) and valves are used to throttle the steam to the required pressure, either to 550 kPa (gauge) for the boilers or 300 kPa (gauge) to 120 kPa (gauge) for the cookers in FB. A proportion of the steam condensate is recovered and reused, and the shortfall is replenished with water from the mains, available at 20°C.

The steam production system consists of five boilers producing steam at 800 kPa (gauge), their information is shown in Table 3.8. Boilers 1 and 2 are operating efficiently and able to produce steam at the rates they were designed for, boilers 3 to 5 are not able to produce steam at the designed rates. A maximum of 38 metric tons/h of steam can be produced, with coal being the major energy source and one boiler consuming HFO.

Table 3.8: The boilers used in the central steam system that provides steam for Factory B.

Parameter	Boiler 1	Boiler 2	Boiler 3	Boiler 4	Boiler 5
Manufacturer			John Thompson		
Fuel used	Coal	HFO	Coal	Coal	Coal
Achievable capacity (tonne/h)	16	10	4	4	4

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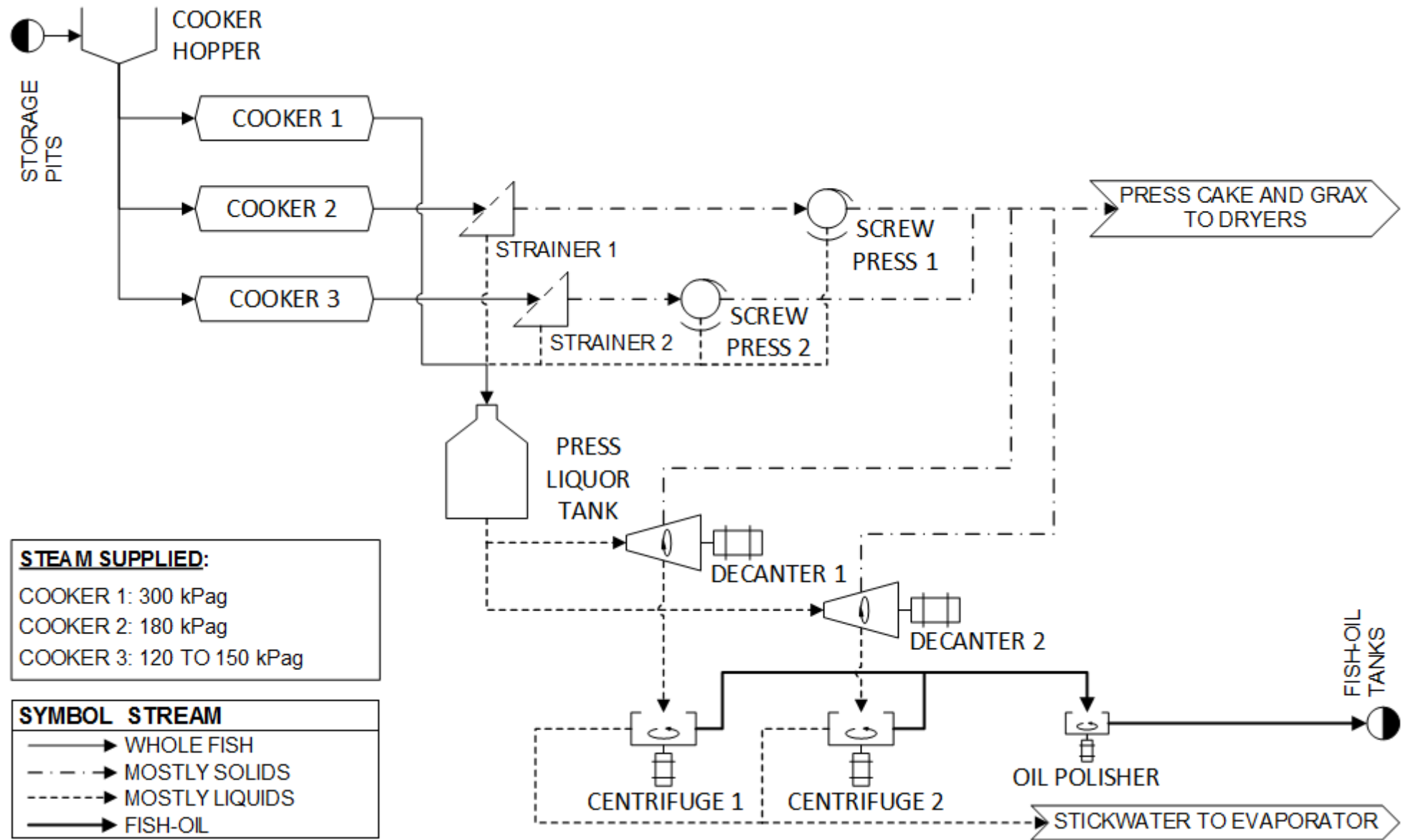
Figure 3-8: Process flow diagram for the wet section of Factory B.
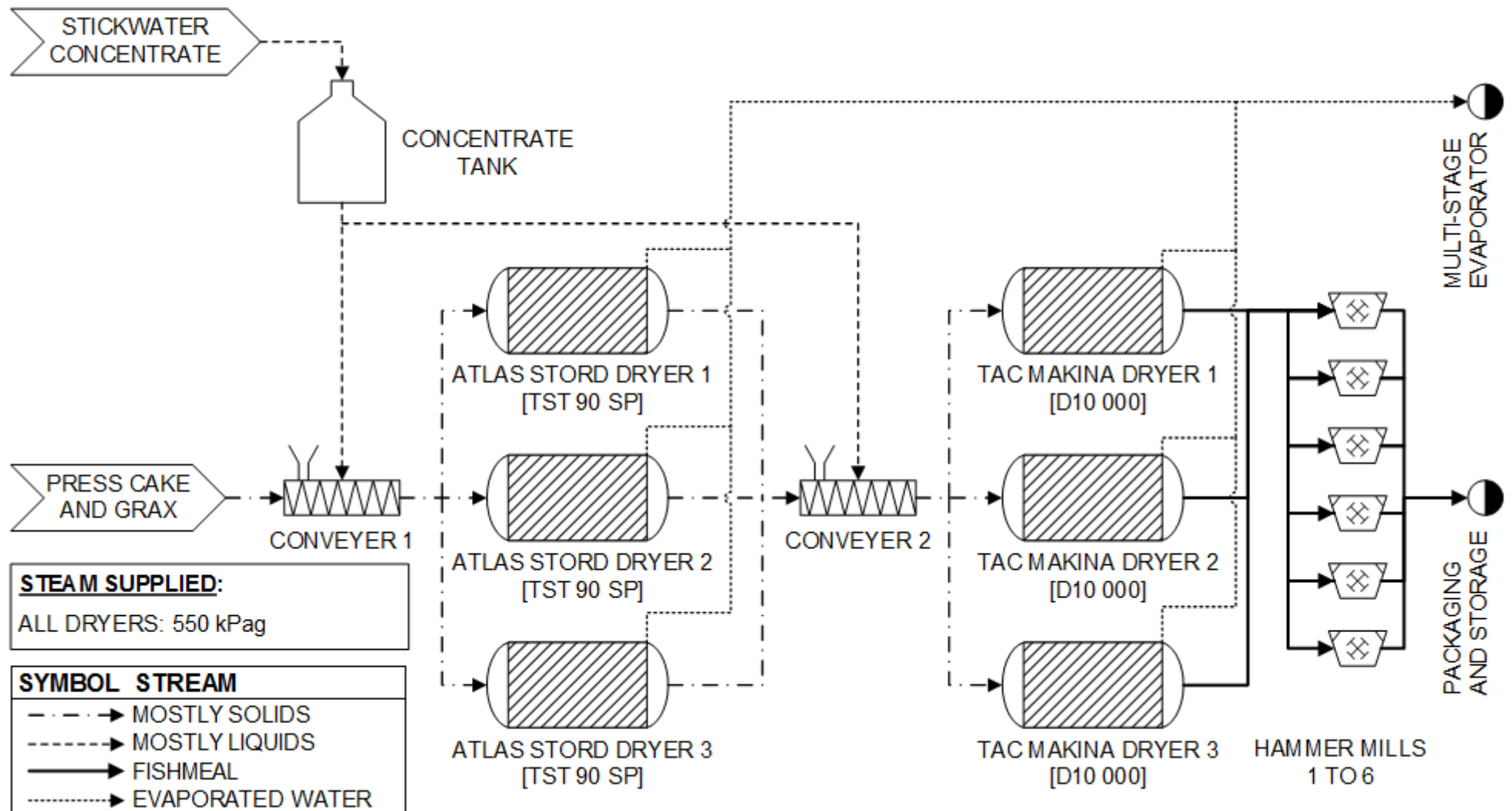
Figure 3-9: Process flow diagram of the dry section of Factory B.

3.5.3 Analysis of Factory B production data

Raw materials processed at Factory B

Production at FB peaks during autumn and the start of winter, when most fish are landed due to largest catches occurring during this period. The amounts of raw material consumed by FB during 2016 are shown in Figure 3-10. During the 2016 production period the greatest mass of raw materials (mostly whole fish) were processed during April to June. Large fish catches are typical during this time of the year, however, catches could differ greatly between years, depending amongst others on the weather and the migration patterns of the fish.
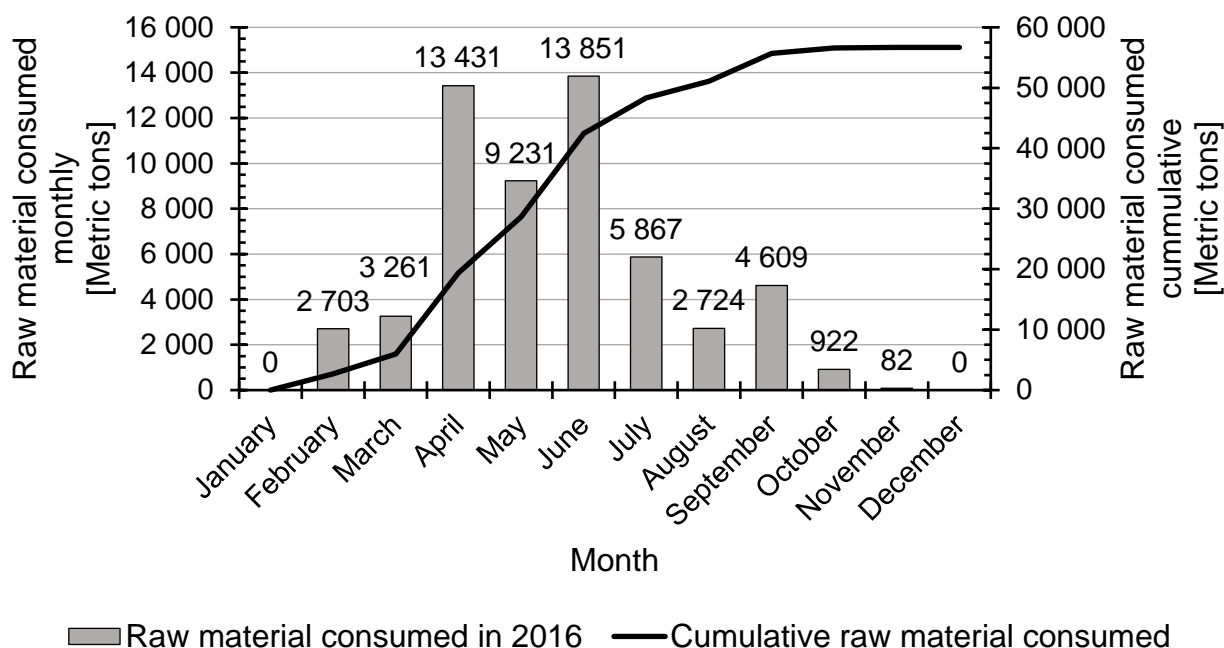
Figure 3-10: Monthly and cumulative amounts of raw material consumed at Factory B during the 2016 production period.

The type of raw material processed was indicated on all the production shift reports, and the contribution of each raw material to the total production during 2016 is shown in Figure 3-11. The raw materials processed per production run were classified as red-eye, anchovy, a mixture of fish species, and cannery offal. Anchovy was processed more than red-eye, and comprised a significant fraction of the raw material when mixed species were processed, according to factory personnel.
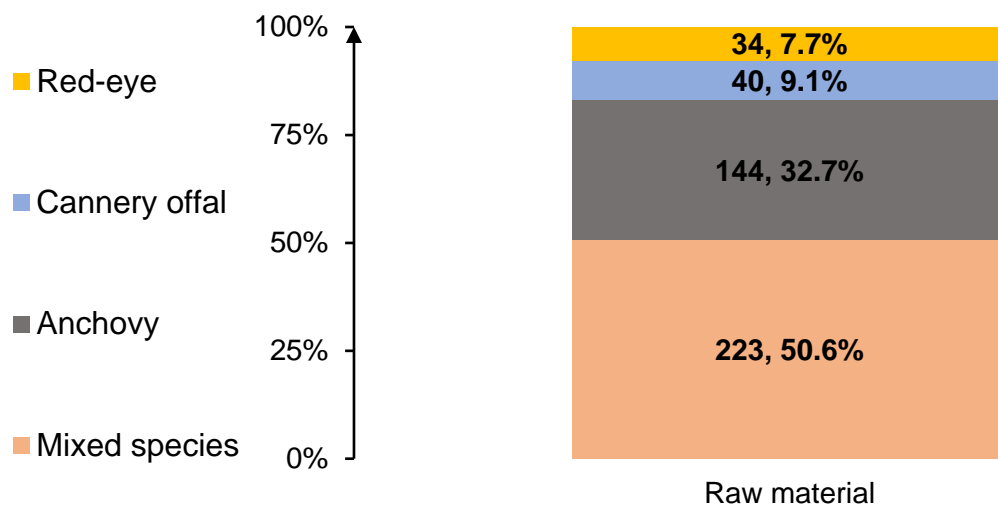
Figure 3-11: Types of raw material processed by Factory B during the 2016 production period. The values indicate the number of shifts that processed a specific raw material, followed by the percentage it comprises of the total number of shifts.

Since fish catches are greatly dependent on the behaviour of the fish shoals, the type of fish landed at FB during the production period also changes throughout the year. The type of raw material processed per shift during the 2016 production period are shown in Figure 3-12. The availability of different raw materials during 2016 was similar to general trends reported by factory personnel: anchovy was available until roughly August and red-eye was available at the start of the season, but not simultaneously with anchovy.
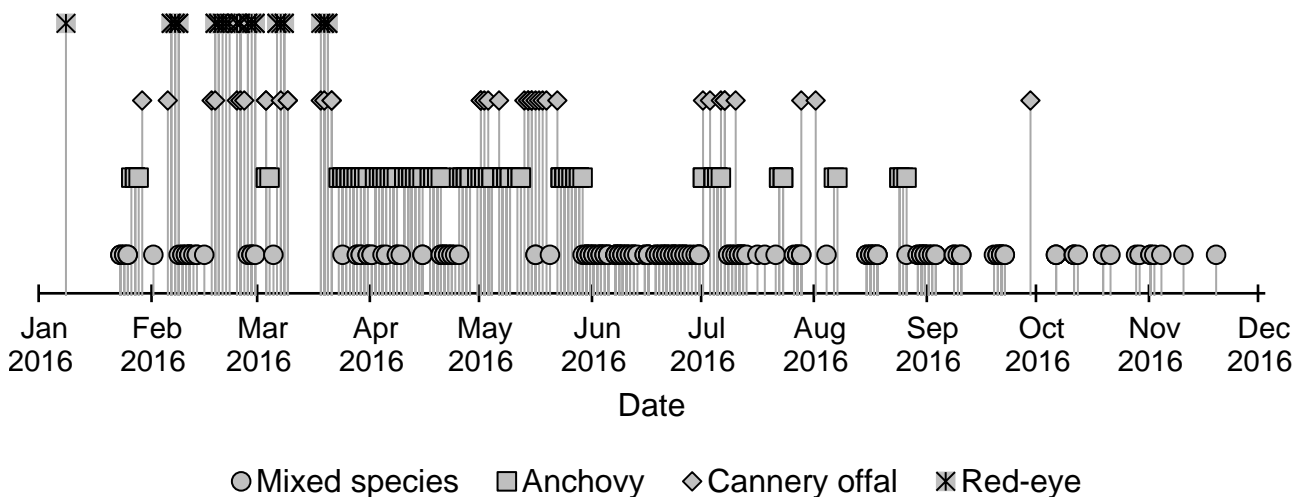
Figure 3-12: Type of raw material processed by Factory B during the 2016 production period.

Raw material specific production information

Since the physical properties are different for each type of raw material, average values of production indicators were calculated for each raw material type, to determine whether the type of raw material processed had an effect on the operation of the plant. Raw-material-specific production information is shown in Table 3.9. The mean production times were similar at approximately 7 hours, except for cannery offal with a mean production time of 5.4 hours, which was processed faster due to the smaller

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amounts thereof. Anchovy were processed in the greatest amounts at 199 metric tons, and at the highest raw material feed rate of 31.81 metric tons/h. The highest yield for fishmeal and fish oil was achieved when processing cannery offal at 29% and 4.2% respectively, however, these values were calculated for a smaller number of data points compared to the other materials Generally, the average yields obtained at FB per raw material were higher than the average yield of 23% for South Africa (de Koning, 2005).

Table 3.9: Raw material specific production information for Factory B per production shift.

	Red-eye	Anchovy	Mixed	Cannery offal
Number of entries	24	94	103	18
Average production time [hour]	7.2	6.9	7.0	5.4
Average raw material processed [metric tons]	150	199	185	84
Average feed rate [metric ton/h]	20.9	31.81	27.5	15.8
Average fishmeal produced [metric ton]	35.8	43.2	46.5	18.8
Average fish oil produced [metric ton]	3.7	5.5	3.9	2.9
Average fishmeal yield	25.0%	24.5%	28.6%	29.0%
Average fish oil yield	2.7%	3.4%	2.4%	4.2%
Average fishmeal moisture content	7.1%	7.5%	8.1%	6.7%
Average meal production rate [Metric ton/h]	4.8	6.7	6.8	3.8

Overall production information

Both the shift and daily data sets contained production data for FB during 2016, and should be equal if the amounts per shift are added together for each day. However, the daily and total values for these two data sets could not be reconciled, therefore, average values of production indicators were calculated for both data sets and are shown in Table 3.10.

Values of the same indicator differed greatly between the shift and daily data sets, and the data showed large fluctuations for all indicators calculated, irrespective of the data set used. The mass flow rate of fishmeal was calculated as 6.3 ton/h and 4.3 ton/h for the shift and daily data sets respectively, further the mass flow rate showed large variations with the standard deviation approximately 48% of the mean value in both cases. The mass flow rate of the raw material was calculated as 27.7 ton/h and 18.3 ton/h for the shift and daily data sets respectively. The shift data set showed greater variation for the raw material flow rate than the daily data set, with its standard deviation approximately 62% of the mean, while that of the daily data set was 53% of the mean. The average fishmeal yield was calculated as 26.7% and 24.5% for the shift and daily data sets respectively, which is higher than the average fishmeal yield in South Africa of 23% (de Koning, 2005). However, the fishmeal yield did

show large variations with the standard deviation of the shift data set approximately 48% of its mean, while the standard deviation of the daily data set was approximately 27% of its mean value.

Table 3.10: Average production values for Factory B during 2016, calculated from shift and daily data sets. The shift data set contained 239 entries for fishmeal and fish oil, while the daily data set contained 75 entries for fishmeal and 59 for fish oil.

Parameter	Source	Value (average)	Standard deviation	Unit
Fishmeal production rate	Shift worksheet	6.3	± 3.0	Metric ton/h
	Daily productions	4.3	± 2.1	Metric ton/h
Fish oil production rate	Shift worksheet	728	± 1 549	kg/h
	Daily productions	362	± 253	kg/h
Raw material feed rate	Shift worksheet	27.7	± 17.1	Metric ton/h
	Daily productions	18.3	± 9.7	Metric ton/h
Fishmeal yield	Shift worksheet	26.7	± 12.9	%
	Daily productions	24.5	± 6.6	%
Fish oil yield	Shift worksheet	3.0	± 5.4	%
	Daily productions	2.0	± 1.3	%

3.5.4 Factory B base case process simulation

Fishmeal and fish oil production process simulation

The FMFO production process of FB was simplified to only include units that change the temperature or composition of streams, that would be of interest to the SHIP study (Chapter Four) and that sufficient information was available for. This simplified process formed the base case and is shown in Figure 3-13. The cookers were combined into a single unit since no information on individual cooker use was available. The three dryers in each drying stage are identical, for the purposes of this study, and were combined into a single unit for each drying stage. The multi-effect evaporator at FB is operated as a separate unit that receives sufficient energy from the water evaporated in the fishmeal dryers, therefore, it was not considered in the base case.

The mass and energy balance calculations for the base case process of FB were automated using spreadsheets in Microsoft Excel 2016, this enabled simulating the effects that changes to the base case process had on the energy requirements. Due to a lack of information not all temperatures could

be calculated, therefore, there are empty values in Figure 3-13. The characterisation of the base case process: temperatures, stream composition and mass flow rates are discussed in the remainder of this section. Images of the base case process simulation of FB in Microsoft Excel are shown in Appendix C.

Raw material composition and feed rate in the base case process

Although mixed raw material was consumed the most, anchovy was selected as the raw material to be used in the base case process, since it is the most consumed single species and the composition of the mixed material was unknown. Furthermore, FB based their own mass balance and process development on anchovy.

The raw material stream of the base case process of FB was characterised with a composition provided by the factory, and the mass flow rate calculated form the production data collected. FB provided composition data for various raw materials in terms of the dry matter (which includes proteins and insoluble material e.g. ash) and oil content mass fractions, and the water content was calculated as the remainder. The raw material composition for anchovy used in the base case process was 16% dry matter, 7% oil and 77% water. The raw material mass flow rate defined for the base case process was 31 810 kg/h; the average value calculated for anchovy from the shift data.

Typical operating temperatures for Factory B

Various sources were consulted to characterise all the operating temperatures of the base case process, which are shown in Table 3.11. The temperatures and stream numbers are the same as those shown in Figure 3-13. The term 'verbal' refers to data which were obtained from personal communications with technical staff at FB, while 'data' refers to values which were calculated from actual factory data, and 'theoretical' refers to values calculated from a combination of literature values and by making certain assumptions.

The outlet temperatures of the cookers and dryers were determined from the PLC data obtained. Due to the high number of data points, it was not possible to analyse all the data, therefore, May 2016 and May 2017 were selected for analysis, based on the fact that the month of May was when the most anchovy were processed and FB stated that May is usually a peak production month. Data for two years were selected to see if there were any changes in the operating conditions over the course of a year, and the average values for these months were used in the base case process. The outlet temperatures of the cookers and dryers, recorded by the PLC's, are shown as graphs in Appendix C.

FB cooked raw materials to 98.5°C, which is within the typical range of 95°C to 100°C as stated by the FAO (1986). The high cooking temperature, relative to the 85°C to 90°C stated by the IFFO

(2016), was most likely to ensure a favourable temperature for oil and stickwater separation during centrifugation (Bergé, 2016).

The outlet temperature of the first drying stage was equal to the literature value of 90°C (FAO, 1986). The outlet temperature of the second drying stage at 97°C was higher, this was most likely required to ensure the required low water content of the fishmeal exiting the second drying stage was achieved.

Table 3.11: Typical operating temperatures and their sources defined for the base case process of Factory B.

Location	Stream	Temperature [°C]	Source
Inlet to cookers	1	22.0	Verbal
Outlet of cookers	2	98.5	Data
Stickwater before evaporation	8	65.0	Verbal
Stickwater concentrate in tank	10	50.0	Verbal
Press cake and grax before adding concentrate	6	83.3	Theoretical
Inlet to first drying stage	6 and 10	77.5	Verbal
Outlet of first drying stage	11	90.0	Data
Inlet to second drying stage	11 and 13	83.5	Theoretical
Outlet of second drying stage	14	97.0	Data

The only theoretical value affecting the base case results, was the inlet temperature to the second drying stage. It was assumed that heat losses between the drying stages were negligible; the temperature of the combined solids and stickwater concentrate stream was then calculated using Equation 3.1. The assumption seems reasonable since the materials only travel a short distance between drying stages, and at steady state the equipment would have been heated up, reducing heat losses.

Equipment operating parameters

The available literature (FAO, 1986, Windsor, 2001) provide typical compositions of streams for the wet-pressing production process, however, this is either stated as a range or is dependent on the raw material composition, and little information on the performance of unit operations is given. If a raw material with a different composition were to be used it would be very difficult to perform a complete mass balance and determine the product composition accurately with the information available in literature.

FB provided equipment performance parameters that describe the behaviour of different process units (e.g. the separation efficiency or selectivity towards solids), and other operating parameters for a variety of raw materials processed. These parameters are representative of FB and remained relatively constant regardless of raw material processed, as an example the parameters for anchovy are shown in Table 3.12. This information enables calculating the composition of outlet streams

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based on fixed equipment performance parameters and not a fixed inlet composition, thus, the composition of streams and products for raw materials with different compositions to that used in literature can be calculated beforehand. The compositions of streams for the base case process were calculated using the parameters shown in Table 3.12.

Table 3.12: Typical operating parameters for equipment in Factory B, when processing anchovy.

Parameter	Value
Dry matter in press liquor as suspended solids	50%
Dry matter in press liquor as dissolved solids	50%
Press cake dry matter content	40%
Press function: fraction of dry matter entering decanters that exits in the press cake.	60%
Unavailable oil	29%
Fraction of unavailable oil in press cake	75%
Grax dry matter content	30%
Grax oil content	5%
Decanter function: fraction of suspended solids removed by decanter	85%
Fraction of dissolved solids exiting in decanter grax	10%
Separator function: fraction of incoming oil separated from	95%
Stickwater concentrate dry matter content	30%
Recommended water content in fishmeal	10%

Figure 3-13: Illustration of the base case simulation of Factory B. The process was simplified to include only temperature or composition changes.
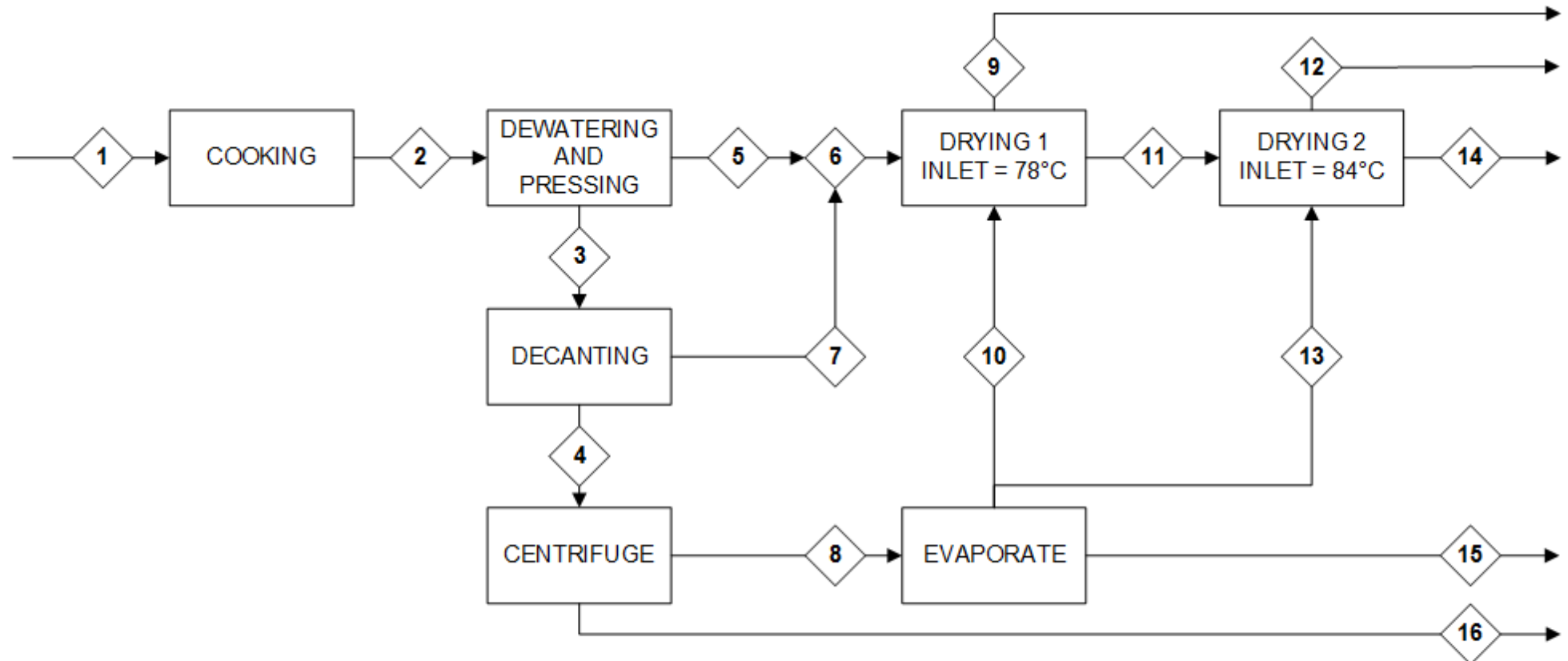
Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
Temperature (°C)	22	99				83		65	90	50	90	97	50	97		
Mass Flow (kg/h)	31810	31810	24176	20952	7634	10858	3223	19443	4082	2138	8913	4062	1425	6276	15880	1509
Composition (Component Mass/Total Mass)																
Dry matter	16%	16%	8%	5%	40%	37%	30%	5%	0%	30%	52%	0%	30%	81%	0%	0%
Oil	7%	7%	7%	8%	6%	6%	5%	0%	0%	2%	8%	0%	2%	11%	0%	100%
Water	77%	77%	84%	87%	54%	57%	65%	94%	100%	68%	40%	100%	68%	7%	100%	0%

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Determining the protein and ash content of streams

FB provided raw material compositions in terms of dry matter, oil and water and the operating parameters (Table 3.12) also catered for only these components. However, to determine the energy required to heat the materials (Equation 3.3), the composition in terms of protein, ash, oil (or fat) and water must be known, therefore, the mass balance needed to be expanded to include the protein and ash content. The water and oil content were known, while the protein and ash content needed to be determined.

It was assumed that the dry matter consisted only out of protein and ash, furthermore, the ratio of protein to ash in the various process streams was assumed to be constant throughout the process. Literature was consulted to obtain the protein and ash content of anchovy-derived fishmeal and the values are reported in Table 3.13. The composition of whole South African anchovy was also included, since FB added the stickwater concentrate back to the solids, and the oil and other effluent streams contained negligible amounts of protein and ash, thus, the ratio of protein and ash for whole fish and fishmeal produced at FB is believed to be similar.

Dry matter was assumed to consist out of only protein and ash, then the fractions of protein and ash in dry matter were calculated for each literature source and the results are shown in Table 3.13. The protein and ash content of anchovy-derived fishmeal were relatively constant across the different sources, with the protein content approximately 65% and the oil content approximately 15%, while the protein and oil content of whole South African anchovy were 17% and 3% respectively. The dry matter in whole anchovy had a higher protein fraction than that of the anchovy derived fishmeal, however, there was still a good comparison between the values.

The average of the calculated protein and ash fractions in dry matter for the five sources in Table 3.13 were taken and used in the base case process. It was assumed that the dry matter of all anchovy processed by FB consisted of 82% protein and 18% ash. The protein and ash content of streams in the base case processes was calculated by multiplying the dry matter content with the protein or ash factions.

Table 3.13: Protein and ash content of anchovy and anchovy derived fishmeal, the share of protein and ash to the total protein and ash content was calculated and the average values presented.

Description	Source	Protein content	Ash content	Protein fraction	Ash fraction
Anchovy fishmeal	Cho and Kim (2011)	64.6%	15.0%	81.2%	18.8%
South African anchovy/pilchard fishmeal	Barlow and Windsor (1984)	66.1%	14.1% [1]	82.4%	17.6%
Anchovy fishmeal	Barlow and Windsor (1984)	65.4%	15.4%	80.9%	19.1%
Mostly whole anchoveta fishmeal	FAO (1986)	65.0%	16.0%	80.2%	19.8%
South African anchovy (whole)	FAO (1986)	17.0%	3.0%	85.0%	15.0%
Mean ± standard deviation				82.0% ± 1.7%	18.0% ± 1.7%

Expanding the mass balance to include protein and ash components

Using the calculated protein and ash fractions in the dry matter of anchovy, the ash and protein content of process streams in FB were calculated and the results for the raw material and fishmeal product streams are shown in Table 3.14. A compositional analysis representative of approximately 450 metric tons of fishmeal produced at FB in January 2017, is also shown in Table 3.14.

The simulation of the base case process in Microsoft Excel 2016 was capable of accurately calculating the composition of the actual fishmeal product stream, based only on the composition of the entering stream and equipment operating parameters. The operating parameters provided by FB originally stated a fishmeal water content of 10%, which was subsequently updated to 7.47%, (which is the actual average moisture value obtained from assaying the anchovy fishmeal produced by FB). There was an excellent comparison between the calculated fishmeal composition and the actual composition obtained via analysis for FB. This increased confidence that the base case simulation was representative of the operations of FB, especially considering that, apart from updating the water content, the actual fishmeal composition values were not used in the base case process mass balance calculations.

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[1] The ash content was calculated by subtracting the protein, fat and water content from the total, since these values were provided.

Table 3.14: Composition of raw material and fishmeal product for the base case process of Factory B, and the composition of fishmeal produced in January 2017.

Component	Calculated raw material composition	Calculated fishmeal composition	Analysed fishmeal composition
Protein	13.1%	66.5%	66.4%
Ash	2.90%	14.6%	14.2%
Oil	7.00%	11.4%	10.4%
Water	77.0%	7.50%	8.10%

Steam production system simulation

The steam production system of FB was simulated with Aspen Plus V8.8 process simulation software, and the simulation flowsheet is shown in Figure 3-14. The cooking and, first and second drying stages were simulated as heat sinks, with the outlet conditions of the condensate specified and the mass flow manipulated to obtain the required amount of energy transferred to each unit. The boilers produced steam at 800 kPa (gauge), which was throttled with valves to 180 kPa (gauge) for cooking and 550 kPa (gauge) for the dryers. The simulation was constructed as follows:

i. Water from the mains (FRESH) at 20°C, joins the recovered condensate (RETURN) in the hot well tank (HWT). The mass flow rate of fresh water was manipulated to ensure sufficient steam was generated and the heating duty required by the second drying stage (DRYING2) was obtained.

ii. Water from the HWT is pumped at 800 kPa (gauge) to the boiler, where it generates saturated steam at 800 kPa (gauge).

iii. A design specification was used to manipulate the fraction of steam diverted (SPLIT1) to the cooking stage (COOKING) to ensure the required amount of heat was transferred. The medium pressure level of steam required by the cookers, at 180 kPa (gauge), was used in the simulation, since any of the cookers could be used and it was not possible to determine what specific cooker was being used at any one point in time. Steam was throttled from 800 kPa (gauge) to 180 kPa (gauge) with a valve (VALVE1).

iv. A design specification was used to manipulate the fraction of steam diverted (SPLIT2) to the first drying stage (DRYING1) to ensure the required amount of heat was transferred. Steam was throttled from 800 kPa (gauge) to 550 kPa (gauge) for the two drying stages with valves (VALVE2 and VALVE3). The remaining steam flow was sufficient to satisfy the heat requirements of the second drying stage (DRYING2), due to the design specification which manipulated the flow of fresh water and, consequently, the steam mass flow rate.

v. Steam traps after the cooking and drying equipment exit to 50 kPa (gauge); this causes a fraction of the steam to be flashed, which joins the evaporate from the fishmeal dryers sent to

the multi-effect evaporator. The steam traps were simulated as adiabatic flash units with a pressure of 50 kPa (gauge), which resulted in outlet conditions similar to those stated by technical personnel at FB. The vapour streams of the flash units were combined (STEAM) into a single stream to calculate the properties of the stream sent to the multiple-effect evaporator (TOWHP).

vi. The liquid streams from the flash units were combined (CONDENS) into a single stream. FB used some of the condensate for hot water purposes, and there were other steam losses in the system, thus, a significant amount of condensate was not recovered. The average condensate recovery during 2016 was 28.3%, calculated from the actual boiler data. A condensate recovery of 28.3%, calculated as the mass flow rate of the RETURN stream divided by the mass flow rate of the stream exiting the boiler, was obtained by manipulating the split fraction of the LOSSES block. The total amount of condensate lost was simulated with a single stream (LOST).
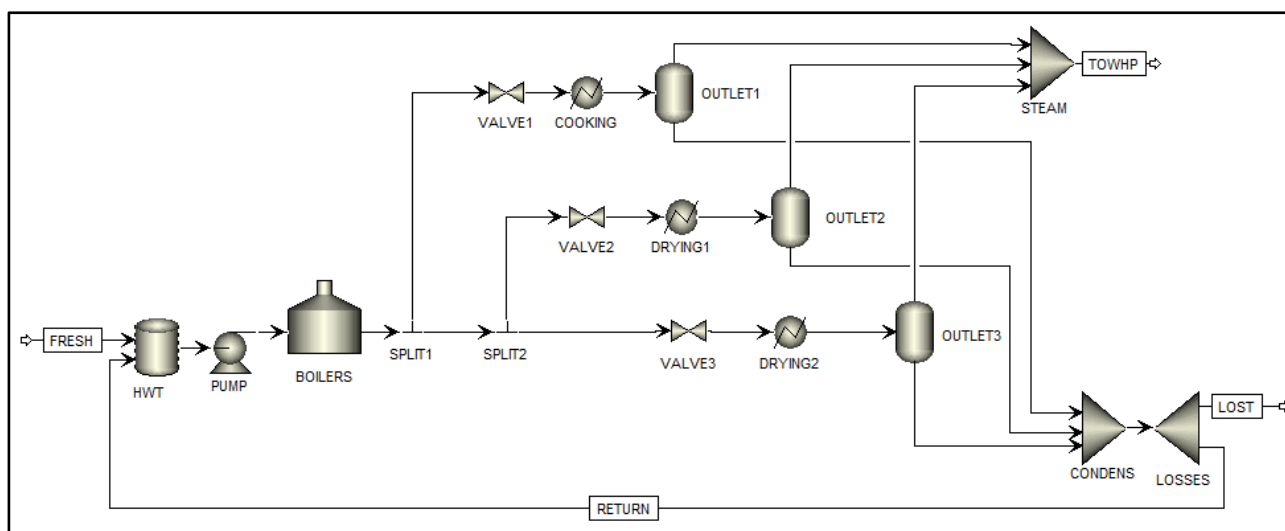
Figure 3-14: Aspen Plus V8.8 simulation flowsheet of Factory B steam production system and steam users.

3.5.5 Factory B base case process energy and fuel requirements

Energy required for cooking and drying

The heating and evaporation of water was responsible for most of the energy required by FB, since it was present in the greatest amount in the cookers, and required evaporation in the dryers. The energy required to cook and dry the raw material with base case process mass flow rates is shown in Figure 3-15. Cooking required 8 864 MJ/h and the drying energy requirements were divided between the first and second drying stages, which required 9 846 MJ/h and 9 606 MJ/h respectively. It was assumed that water evaporated at the outlet temperature of each dryer, with the removal of steam to the evaporator reducing the pressure inside of the dryers.
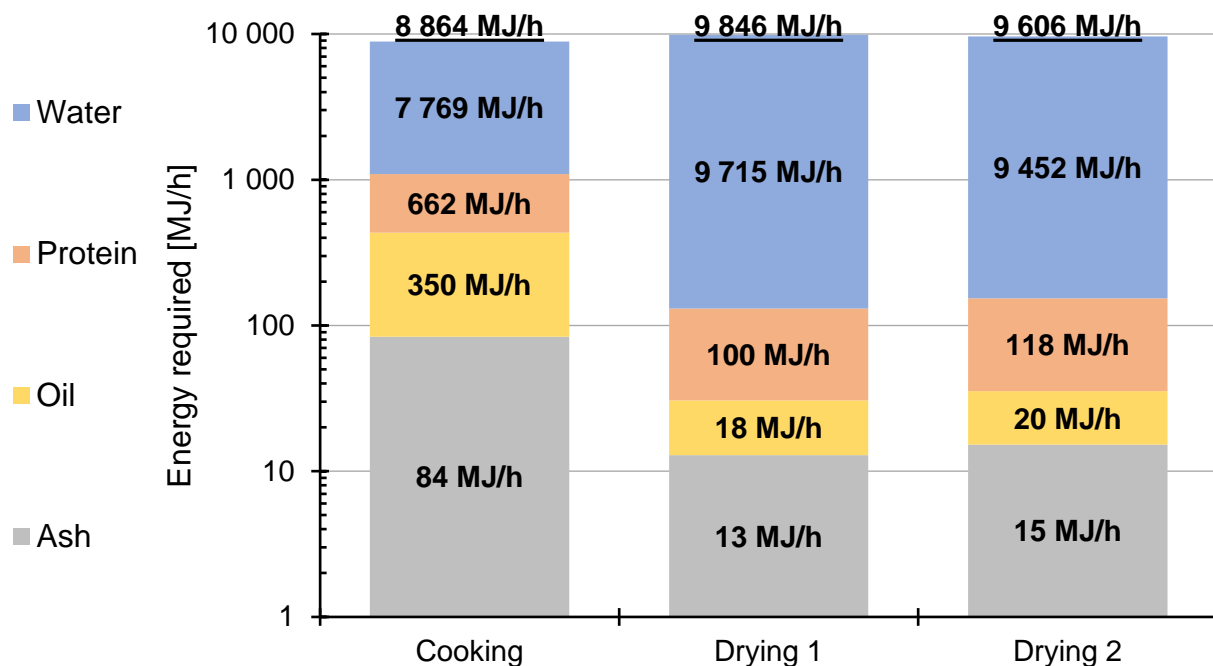
Figure 3-15: Energy required to cook and dry 31 810 kg/h of raw material in Factory B. The average outlet temperatures were: 98.5°C for cooking, 90.0°C for drying 1 and 97.0°C for drying 2.

Equipment efficiencies

The theoretical amounts of energy required to cook and dry the raw material have been established, however, these values are the minimum required energy due to heat transfer inefficiencies in the equipment. FB could not provide any values for equipment heat transfer efficiencies, therefore, the efficiencies needed to be calculated theoretically. Two methods were proposed: 1) with the energy transferred to the materials, and the inlet and outlet conditions of the steam known, if the steam mass flow rate was known, the energy transferred to the equipment and the efficiency could be calculated; 2) the equipment manufacturers could be consulted to obtain the heat transfer efficiencies.

The first method did not yield results, due to a lack of information on the steam flow to individual equipment units. The boiler shift data contained only information relevant to the steam production system (such as the total amounts of steam produced and fuel consumed), not the destination of the steam. Neither the total steam flow to FB, nor the steam flow to individual equipment units were measured, without which the equipment efficiencies could not be calculated.

No information about the heat transfer efficiencies of fishmeal production equipment could be obtained from either Atlas Stord (fishmeal dryer manufacturer) or taç makina, (cooker and fishmeal dryer manufacturer), therefore, the second method also did not yield useful results. Furthermore, a single efficiency value obtained from a manufacturer would not be accurate since it is affected by the equipment age, the operating regime and, frequency of maintenance and cleaning.

Since no reliable heat transfer efficiency values could be obtained, the energy required by cookers and dryers for the base case process was defined as the energy required to cook and dry the raw

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material (Figure 3-15), i.e. the heat transfer efficiency was taken as 100%. Any inefficiency in the actual equipment would result in a greater amount of energy required from steam and even more fuel consumed due to the compound effect of the inefficiencies of subsequent equipment pieces. Efficiencies of less than the 100% assumed in the base case process, would be favourable for the economic viability of the ST systems investigated in Chapter Four, since the heat required from the less efficient steam system would be provided by solar heat instead and the fuel savings would increase with greater inefficiencies.

Steam and fuel requirements

Under base case process conditions, cooking and drying of raw material required 13 192 kg/h of steam in total. The simulation of the steam production system in Aspen Plus V8.8 was used to calculate the mass flow rate of steam to provide 8 864 MJ/h to the cooking stage, and 9 846 MJ/h and 9 606 MJ/h, to drying stages 1 and 2 respectively, and the results are shown in Figure 3-16. The evaporation of water in the drying stages results in greater energy demand and, thus, the two drying stages consumed the largest amounts of steam.
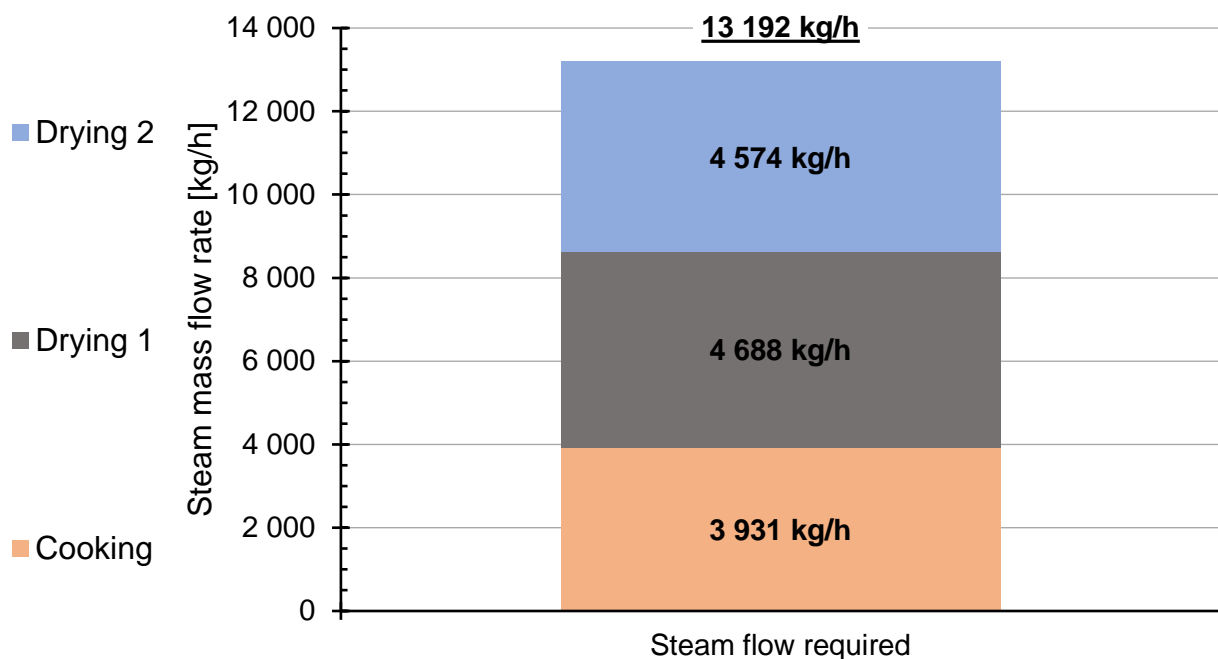
Figure 3-16: Required mass flow rate of steam for base case process of Factory B.

To determine the fuel requirements for the base case process, it was assumed that the most efficient boilers, that consume the cheapest fuel, would be used first, followed by the less efficient. Boiler 1 (Table 3.8) has a capacity of 16 000 kg/h and can produce all the steam required by the base case process, thus, coal was assumed to be the only fuel source.

The boiler data were investigated and it was found that for 75 data entries, steam was produced using only coal. Using only these data, which are shown in Appendix C, it was calculated that an average

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of 6.35 kg of steam was produced for every kg of coal consumed. To produce 13 192 kg/h of steam would require 2 078 kg/h of coal, which characterises the coal usage for the base case process.

To enable accurately calculating the coal requirements, when deviating from base case conditions, a method based on the energy requirements of the boiler and not the steam mass flow rate was developed. For the base case, the boiler required 34 728 MJ/h and consumed 2 078 kg/h of coal, the quotient of these numbers show that 59.8 grams of coal was required to provide 1 MJ of energy to the boiler, this value was subsequently used to determine mass flow rate of coal required. A gross calorific value of 27.5 MJ/kg for coal, provided by John Thompson boilers (John Thompson, 2017) was used where necessary.

3.5.6 Factory B yearly production schedule

The production schedule of FB for 2016 was determined from actual production data, and expressed in terms of the fraction of raw materials consumed during a specific month to the total amount of raw materials consumed during the year. According to the complete daily production data set, FB consumed 56 681 metric tons of fish during 2016, which falls within the allowable annual quota of the facility. The actual production schedule of FB during 2016 is shown in Figure 3-17, as stated by FB personnel: the months of April, May and June were the peak months, with little to no materials being processed during January, November and December.
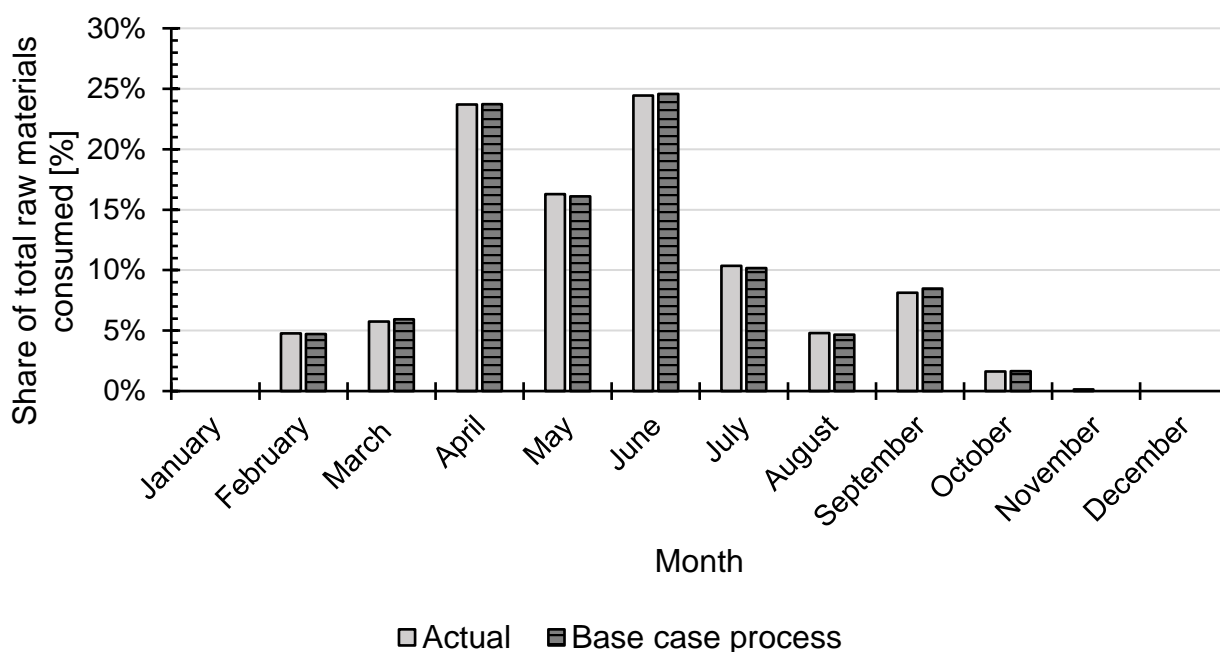
Figure 3-17: Raw material consumed by Factory B during 2016 in reality and for the schedule defined for the base case process

To characterise the yearly schedule for the base case process, the actual monthly shares of fish consumed were multiplied by the quota of FB to obtain the quantity of raw material consumed per

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month. The monthly quantities were divided by the base case feed rate of 31.81 metric tons/h to calculate the production hours required per month to process the raw material. Production shifts were assumed to be 8 hours long, with shifts starting at 07:00 am, 03:00 pm and 11:00 pm (obtained from shift worksheet data). The shift duration appears reasonable when considering that the actual production times, were slightly less than 8 hours (Table 3.9), which is expected since some time is lost during start-up, where materials are recirculated until the required temperatures are reached.

The number of possible shifts per month were calculated by multiplying the number of days with 24 hours, since the factory operates 24 hours a day, and dividing this by 8 hours to obtain the number of shifts. The number of shifts required to process the raw materials were calculated by dividing the number of required hours by 8. It was assumed that the production and non-production (cleaning and maintenance) shifts are distributed evenly throughout a month, consequently, the shift schedule per month could be created. The monthly shares of raw materials consumed for the base case process schedule is also shown in Figure 3-17; there was a good comparison between the actual values and those of the schedule defined for the base case process.

3.6 Base case process results for Factory A and Factory B

3.6.1 Comparison of Factory A and Factory B

The production amounts and resource requirements of the factories according to the base case processes are presented in this section. To summarise the base case process results, the factories are compared in terms of: hourly production rates, production for a year-long period and on the basis of 1 000 kg of fishmeal produced.

<u>Production time based comparison</u>

Due to multiple production lines (three in parallel) and process units with larger capacities, FB is capable of greater production rates than FA. The hourly production rates and resource requirements for one hour of production at both factories are shown in Table 3.15. Based on the collected production data, FB typically produces fishmeal at a rate of 6 276 kg/h (1 509 kg/h for fish oil) compared to the 307 kg/h for FA, and consumes raw material at a rate of 31 810 kg/h compared to the 1 327 kg/h of FA.

Based on the base case process results, FA better utilises the energy released from fossil fuel combustion than FB. FA utilised 72% of the energy released from combusting HFO by effectively transferring 2 837 MJ of the 3 964 MJ released to the raw material, while FB only utilised 50% of the energy released from combusting coal by effectively transferring 28 316 MJ of the 57 132 MJ released to the raw material. It is believed that the lower efficiency of FB is due to a less efficient

steam production system compared to FA, and low amounts of condensate being recovered within the production process, which leads to high energy losses.

Table 3.15: Summary of resource requirements for one hour of production at Factory A and Factory B according to the base case processes.

Parameter	Unit	Factory A	Factory B
Fishmeal produced	kg	307	6 276
Raw material consumed	kg	1 327	31 810
Water removed	kg	1 019	24 025
Fuel consumed	FA [litre] FB [kg]	100	2 078
Live steam produced	kg	1 460	13 192
Energy transferred to fish material	MJ	2 837	28 316
Energy from fossil fuels	MJ	3 964	57 132

Annual production comparison

FA is a small factory that processes mostly fish-processing by-products in relatively low amounts, while FB processes whole pelagic fish species in considerably larger amounts than FA. The annual production amounts and resource requirements of the base case processes of both factories for the year 2016 is shown in Table 3.16. Based on the year 2016, FA produced 1 610 tons of fishmeal, and FB produced 11 850 tons of fishmeal (2 849 tons of fish oil), according to their base case processes.

FA processed fish-processing by-products which are more frequently available, albeit in small quantities, the result being that FA produced fishmeal for a large proportion of the year. However, FB consumed freshly landed pelagic species, which are typically available in larger quantities and have to be processed much quicker to avoid spoilage. For 2016, FA had a total production time of 5 240 hours, which was significantly more than the 1 888 hours of FB. However, FB still processed substantially more raw material at 60 057 tons compared to the 6 952 tons processed by FA. FA sustained long production times by continuously obtaining small amounts of raw material, while FB had to quickly process large amounts of raw material leading to a shorter production time and higher production rates (Table 3.15).

Table 3.16: Summary of annual resource requirements of Factory A and Factory B according to the base case processes based on the year 2016.

Parameter	Unit	Factory A	Factory B
Total production time	Hours	5 240	1 888
Fishmeal produced	Tons	1 610	11 850
Raw material consumed	Tons	6 952	60 057
Fuel consumed	FA [m^3] FB [Tons]	525.0	3 922
Live steam produced	Tons	7 652	24 907

Production mass based comparison

FA required significantly more time to produce 1 000 kg of fishmeal than FB, however, FA achieved a higher fishmeal yield. The resource requirements to produce 1 000 kg of fishmeal at both factories are shown in Table 3.17. FA processed 4 318 kg of raw material within 195 minutes to produce 1 000 kg of fishmeal, while FB processed 5 068 kg of raw material in 10 minutes to produce the same mass of fishmeal. According to the results of the base case processes, FA achieved a fishmeal yield of 23% compared to the 20% of FB. The yield of FA was expected to be higher, since its raw material had a higher protein content (15.5% compared to 13.1% for FB) and FA only produced fishmeal as product, while a portion of the raw material of FB resulted in the production of fish oil.

Per unit mass of fishmeal produced, FA required more energy and consequently more steam than FB. FB required 2 102 kg of steam to produce 1 000 kg of fishmeal, while FA required more than double this amount at 4 753 kg. The difference is attributed to the fact that FB recovers waste heat from the drying stage by using the evaporate from the fishmeal dryers to concentrate the stickwater in the multiple effect evaporators, which are also more efficient drying units. The triple-effect evaporator of FB consumed 0.51 kg of steam for every kg of water evaporated in the base case process, which is slightly higher than the 0.37 to 0.45 kg steam/kg evaporate range stated in literature (Table 3.1).

Table 3.17: Summary of resource requirements to produce 1 000 kg of dry fishmeal at Factory A and Factory B according to the base case processes.

Parameter	Unit	Factory A	Factory B
Fishmeal produced	kg	1 000	1 000
Raw material consumed	kg	4 318	5 068
Water removed	kg	3 318	3 828
Fuel consumed	FA [litre] FB [kg]	326	331
Live steam produced	kg	4 753	2 102
Energy transferred to fish material	MJ	9 235	4 512
Energy from fossil fuels	MJ	12 903	9 103
Time required	Minutes	195	10

3.6.2 Energy consumption of the factories compared to literature

The calculated energy requirements for FA, per unit of raw material processed, were greater than the energy requirements of other FAQ fishmeal factories stated in literature. The energy and resource requirements to process 1 000 kg of raw material at FA and FB under typical conditions are shown in Table 3.18. FA required 2 988 MJ of energy from fuels to process 1 000 kg of raw material, which was greater than the typical maximum of 2 406 MJ required by Peruvian plants producing FAQ fishmeal (Fréon et al., 2017). The greater energy consumption was attributed to the unique production process of FA, that requires all the water to be evaporated in a single-stage dryer unit, and the fact that no heat is recovered from the heated vapour stream exiting the dryer.

The calculated energy requirements of FB were within the range of values obtained from literature for other prime fishmeal plants. The 1 796 MJ required by FB to produce prime fishmeal from 1 000 kg of raw materials is more than the 1 498 MJ reported for a Peruvian plant producing the same class of fishmeal (Fréon et al., 2017), however, it is less than the 1 890 MJ stated by Myrvang et al. (2007). The factories that had the minimum and maximum energy requirements encountered in literature had annual raw material capacities of 155 500 and 130 000 metric tons respectively. Compared to the 60 000-metric ton capacity of FB, capacity is apparently not the sole factor affecting energy efficiency. FB required 415 kg of steam per 1 000 kg raw material, which was less than the 528 kg stated by Myrvang et al. (2007).

Table 3.18: Summary of resource requirements to process 1 000 kg of raw material at Factory A and Factory B according to the base case processes.

Parameter	Unit	Factory A	Factory B
Fishmeal produced	kg	232	197
Raw material consumed	kg	1 000	1 000
Water removed	kg	768	755
Fuel consumed	FA [litre] FB [kg]	76	65
Live steam produced	kg	1 101	415
Energy transferred to fish material	MJ	2 139	890
Energy from fossil fuels	MJ	2 988	1 796
Time required	Minutes	45	2

Due to the good comparison with literature values, the base case process mass and energy balances are assumed correct and capable of realistic prediction of the energy requirements of the two factories. The energy requirements of FA were greater than that of other FAQ plants stated in literature, however, the values were still comparable. The energy requirements of FB were similar to those of other fishmeal plants producing prime fishmeal with the wet-pressing method.

4   Possibilities for solar process heat incorporation in two South African fishmeal factories

This chapter focusses on the feasibility of integrating solar heat into the two fishmeal factories studied. The methodologies developed to identify opportunities for solar heat, and to determine a preliminary value for the required solar thermal collector area are given. Opportunities for heat recovery and solar heat integration are identified, and their effects on the process quantified. A preliminary economic analysis is presented and the economic viability of the different integration scenarios discussed.

4.1 Introduction

Knowledge of the FMFO production process and the typical operating temperatures of STC's were used to identify opportunities for solar process heat and heat recovery within the process. The base case processes are believed to be accurate models of the mass and energy balances of the two factories studied, therefore, they were used to quantify the effect of changing operating conditions.

The aim was to identify opportunities to decrease the energy consumption in the factories by providing a fraction of the energy requirements by way of ST energy, which would result in reduced fossil fuel consumption. For any investigated scenario, if the cost savings derived from reduced fuel consumption were greater than the capital and operational costs required, over the defined lifetime of the project, that integration scenario was taken as being feasible.

4.2 Methodology to quantify the feasibility of integrating solar heat

4.2.1 Identifying opportunities for heat recovery and solar heat

Traditional energy efficiency methods

Two common methods which are employed in energy efficiency studies are pinch analysis, and exergy analysis. While these methods provide valuable information, they were developed specifically for the chemicals processing industry and require detailed information about the process and the different unit operations. Furthermore, these methods assume that the process operates continuously and that heat can be efficiently exchanged between all streams.

These methods are less useful for processes that operate on a semi-continuous basis or batch-wise, in cases where limited information is available on the physical properties of streams or equipment efficiencies, or where it is not possible to exchange heat between certain streams. Furthermore, the intermittent nature of renewable energy sources presents an entirely new challenge, and even state-

of-the-art pinch analysis methods are not capable of adequately describing the inclusion of renewable energy, especially solar energy, within an industrial process (Atkins et al., 2010).

In pinch analysis a stream is defined as any flow which requires heating or cooling, but does not change in composition (Kemp, 2007). The fishmeal production process presents a challenge to pinch analysis, since there are very few, if any, streams that require heating or cooling outside of a unit operation that also changes the stream's physical properties or composition. Therefore, a pinch analysis on a typical fishmeal production process would yield limited results as the analysis only includes a small proportion of the total process streams.

<u>Quick identification method developed for non-standard cases</u>

A systematic approach of identifying opportunities for solar heat in non-standard processes was lacking. In this study, a non-standard process is one where information on stream properties or equipment may be unavailable, or where there are significant constrains on which streams and equipment may be considered for heat exchange. From the perspective of pinch analysis, the FMFO production process is such a non-standard process that presents a challenge to traditional energy efficiency and heat integration methods.

The following three step methodology has been developed in this study to aid in identifying inefficiencies and opportunities for ST heat in non-standard production processes:

  i.   Identify heat sinks within the process.
       - Heat sinks are equipment units that require large amounts of energy. This energy may be delivered in the form of electricity, steam or direct heating by combustion of fossil fuels.
       - Determine the temperature at which heat is required by the unit, the outlet temperature may be a good indication. Determine if the design of the unit requires the heat to be delivered in a certain way, for example if it requires steam to operate.
       - Determine if ST technology could be used to provide heat to the unit, keeping in mind the limitations of the technology and the requirements of the unit.
  ii. Identify streams that could be preheated.
       - Any stream that enters the process at a temperature significantly lower than the operating temperature of the first unit it enters, could benefit from preheating, since energy is required to heat the material to operating temperature.
       - Depending on the specific temperatures it is possible that some, or all, of the required energy could be provided by an external energy source.

- Determine a practical temperature to heat the stream to, then based on the initial and final temperatures identify a ST technology that could supply heat at the required temperature.

iii. Identify any streams that exit the process at a high temperature.

- Any stream that exits the process at an elevated temperature represents a potential waste of energy; an elevated temperature was viewed as being at least 10°C warmer than any stream that needs to be heated.

- Determine if the physical properties of the stream allow heat to be efficiently recovered from it. For example, whether the stream is a fluid stream that could be used in a heat exchanger, or a solid particle stream, where heat exchange is more difficult.

- If the temperature and properties of the stream allow for heat exchange, identify a stream that requires heating and that the inlet temperature is at least 10°C colder than the hot stream. Determine if heat exchange between the two streams is practical based on knowledge of the factory layout.

This method can be performed quickly and only requires a basic understanding of the process and the operating temperatures of unit operations and streams. It is not a comprehensive method such as that described by Muster et al. (2015), and would be inadequate for complex production processes, however, it provides useful information in a preliminary study of solar process heat potential, or for a small facility.

4.2.2 Solar heat integration considerations

There is a conflict between the design philosophies of traditional energy systems and ST energy systems. For traditional energy systems the aim is to operate at the highest possible temperatures to decrease the size of heat exchange equipment. However, for ST systems it is desirable to operate at the lowest possible temperatures to maximise the efficiencies of the ST collectors (Schnitzer et al., 2007). Additionally, when ST systems replace fossil fuel based energy systems it is more advantageous to replace higher temperature operations as the fuel savings incurred are higher (Kalogirou, 2003).

The location and temperature at which solar heat is integrated into an industrial process must be carefully considered (Atkins et al., 2010), furthermore, integration with the existing energy supply must be well-matched with the process (Kalogirou, 2003). The most cost effective, commercialised and mature application of ST technology is the heating of water, both at domestic and industrial levels

(Atkins et al., 2010). Water preheating processes are particularly popular applications of ST as the low operating temperatures allow the collectors to operate more efficiently (Mekhilef et al., 2011).

The variable nature of solar heat supply and process heat demand must be taken into account, as both demand and supply could be unsteady and non-continuous (Atkins et al., 2010). Thermal storage may be used to attenuate the fluctuations between demand and supply, and the storage can be used as an integration point for auxiliary energy sources when solar is not sufficient. ST systems without storage may be much cheaper, however, they will not be feasible when the process requires heat at times when solar radiation is unavailable (Kalogirou, 2003).

4.2.3 Determining the required solar thermal collector area

ST systems are sized in multiple ways, however, the procedures used could follow one of two philosophies: either to provide all the required heat for a specific unit, or to provide a fraction of the required heat. If all the heat is to be provided by solar, variations in solar radiation throughout the year will result in the system being correctly sized only for a short period when solar irradiation is at a minimum, but oversized during the rest of the year, which results in inefficient, costly systems that produce excess heat. Therefore, systems that supply the entire heat demand by solar are not encountered in practice, however, as a theoretical case and to establish the total potential for solar heat (in terms of total solar collector area) in a specific factory this scenario was included in the current study. When only a fraction of the required heat is supplied by solar, the system can be more appropriately sized, however, an auxiliary energy source is required to provide the remaining heat (Atkins et al., 2010).

Methodology to determine a preliminary value for solar thermal collector area

With opportunities for heat recovery and solar heat identified, the simulations of the base case process (in Microsoft Excel 2016 for the fishmeal production process and Aspen Plus V8.8 for the steam production system) were used to calculate the new energy requirements from steam and solar heat. The yearly production schedule was then used for each factory to determine the energy required from the solar process heat system per month, based on the hourly values obtained from the simulation results.

The energy required from a solar process heat system (solar heat demand) to preheat the raw material stream in FA (discussed in section 4.3) is used as an example in this section. The solar heat demand for this application in FA for the months of April to July, is shown in Figure 4-1 [Left], with the units in kilowatt-hours (kWh) as these are commonly used to present solar radiation values.

In this study, the solar irradiation on a location was characterised with a typical meteorological year (TMY), obtained from Meteonorm 7 (V7.1.11.24422). A TMY is a generated dataset that is

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representative of the meteorological events of a location over several years, therefore, it is well suited to planning a solar process heat system (Kalogirou, 2003). Figure 4-1 [Right] shows the monthly solar irradiation at the location of FA, for a plane facing North-East (46° East from North) with a slope of 30°. Since FA is located in the Southern hemisphere, the solar irradiation reaches a minimum during the winter months of May to July.
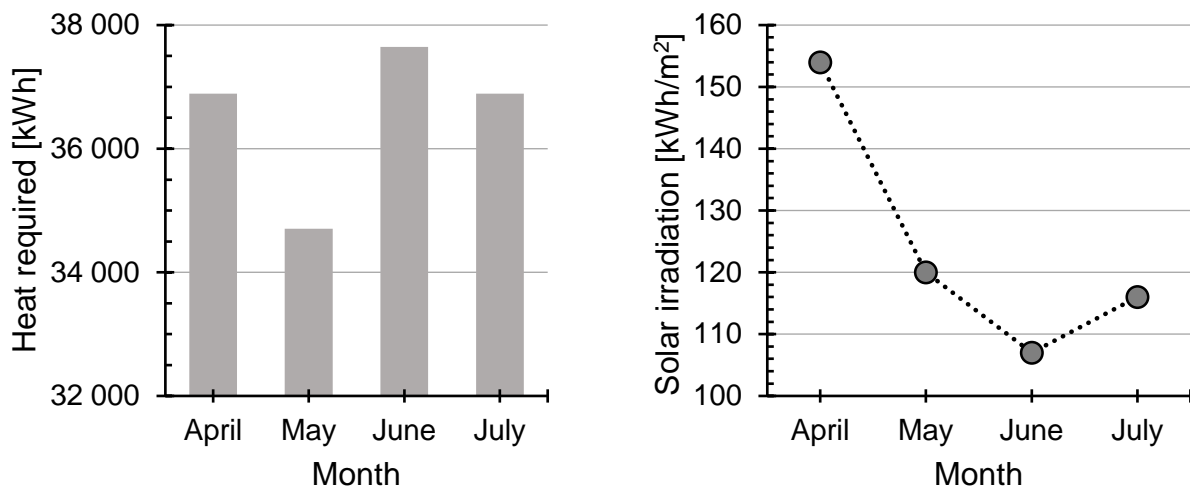
Figure 4-1: [Left] Example of solar heat demand for preheating the raw material stream in Factory A. [Right] Monthly solar irradiation at Factory A's location according to the typical meteorological year obtained from Meteonorm 7.

The total solar heat produced in a month ($Q_{delivered,j}$), is obtained by multiplying the solar irradiation of that month ($G_{total,j}$) with the total ST collector area (A_{total}), however, there are inefficiencies present in the system and an efficiency term (η_{solar}) must be included (see Equation 4.1). An overall efficiency of 45% was used for the solar heat system, which is the same value used by Joubert et al. (2016) to estimate the potential solar collector area that could be used for boiler water preheating in South Africa.

$$Q_{delivered,j}[\text{kWh}] = A_{total}[\text{m}^2] \times G_{total,j}[\text{kWh/m}^2] \times \eta_{solar}[-] \qquad \textbf{Equation 4.1}$$

It is emphasized that Equation 4.1 is a simplification of the actual solar collector output model, and the overall efficiency of 45% is an approximate value, suitable only for preliminary calculations. The useful power output of solar process heat collectors at steady state is adequately described with a model such as Equation 4.2 (BSI, 2013). The type of incident solar irradiation affects the collector output, as the ability to utilise beam (G_b) and diffuse (G_d) irradiation differs between the different collector technologies. The direction of incident irradiation also affects the ability of the collector to utilise it, thus, incidence angle modifiers are included for beam (K_b) and diffuse (K_d) irradiation. Incidence angle modifiers range in value from 0 to 1 and affect the amount of available radiation, depending on the angle at which it hits the solar collector. Due to the temperature difference between

the mean collector fluid temperature and the ambient temperature $(T_m\text{-}T_a)$, there are losses to the environment that are described with the heat loss coefficient (c_1), the temperature dependence of the heat loss coefficient (c_2) is also included.

$$\dot{Q}_{\text{useful}} = A_c[\eta_0 K_b G_b + \eta_o K_d G_d - c_1(T_m - T_a) - c_2(T_m - T_a)^2] \qquad \textbf{Equation 4.2}$$

Following the above discussion, it is clear that the equation used to quantify collector output in this study (Equation 4.1) is a greatly simplified model of reality. However, since the output over an extended period of a month was to be calculated, eliminating instantaneous weather effects, and a constant efficiency was assumed for preliminary calculations, this was deemed acceptable.

To determine the total potential for solar process heat in the fishmeal production process, the collector area that supplied all the required heat throughout the year was calculated. For the example based on the solar heat demand of FA, this scenario is shown in Figure 4-2. Depending on the solar irradiation and heat demand profiles, the system will be properly sized for one month only (June in this case), with excess heat provided during all the other months. As stated by Atkins et al. (2010), this would be an oversized, inefficient, and expensive system, however, it does aid in establishing the total potential for solar process heat in a specific facility.
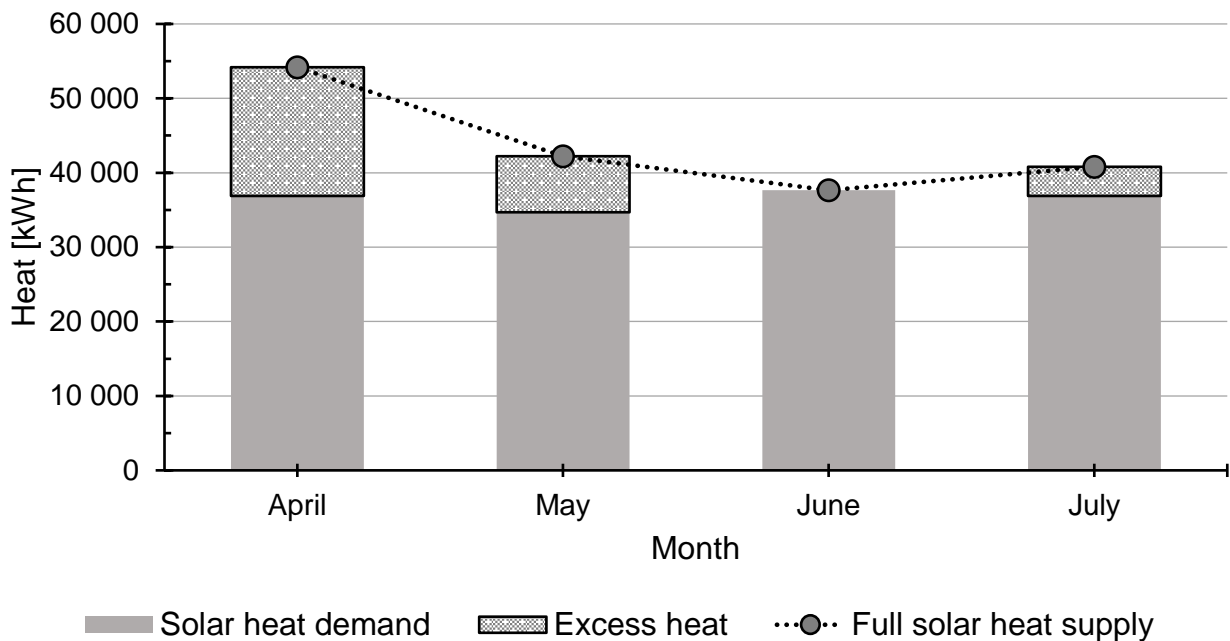
Figure 4-2: Solar heat demand for preheating the raw material stream in Factory A, and heat delivered from a solar heat system with a solar thermal collector area sized to supply the entire heat demand throughout the year.

The alternative is to have a system with a smaller STC area, which provides only a fraction of the required heat. This results in a less expensive system that does not produce as much excess heat. For a preliminary ST study, it is difficult to decide on an appropriate value of the STC area to be used, since the ST industry typically performs detailed simulations of the solar heat supply with multiple

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STC areas to determine a final value. Therefore, two simpler methods of determining the STC area were used in this study: one was developed to minimize the difference between solar heat supply and heat demand for a particular facility over the period of an entire year, and the second to ensure no excess heat is produced during months where production takes place. It was assumed that thermal storage is available that enables all solar heat produced during a particular month to be utilised within that month, but that excess energy cannot be carried over to any of the following months.

For the first method, option A, the overall system efficiency was assumed constant at 45%, with the total collector area as the independent variable. The absolute total difference between heat supply and demand ($Q_{total\ difference}$) was calculated by taking the sum of the absolute values of the difference between solar heat supply ($Q_{solar,j}$) and heat demand ($Q_{demand,j}$) for each month (j) see Equation 4.3. The ST collector area that minimised the absolute total difference was determined by using the GRG Nonlinear solving method in Microsoft Excel 2016.

$$Q_{total\ difference} = \sum_{j=1}^{12} \left| Q_{solar,j} - Q_{demand,j} \right| \qquad \textbf{Equation 4.3}$$

For the second method, option B, the solar collector area was selected to ensure no excess heat is produced during production months. For the example based on the solar heat demand of FA, the resulting heat supply for collector areas calculated with both methods are shown in Figure 4-3. Compared to the scenario where the solar system supplied the entire solar heat demand (Figure 4-2) the excess heat is significantly less for option A. The heat supplied by option B matched the demand during April, and was less than the demand for the other months in this example. With these methods insufficient heat will be provided during certain months and additional heat will be required during these periods. However, the systems obtained are less costly since the total collector areas are smaller, and more usable heat is obtained per ST collector area.
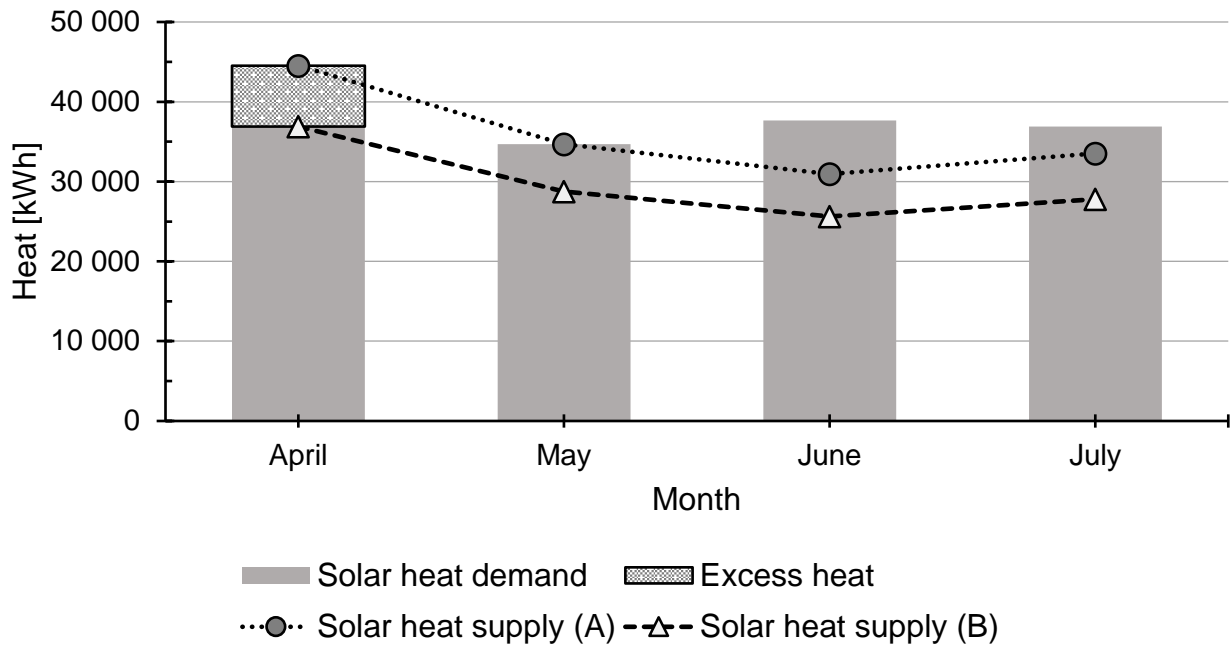
Figure 4-3: Solar heat demand for preheating the raw material stream in Factory A, and solar heat delivered by systems that: A) minimised the difference between solar heat demand and supply, and B) provided no excess heat during months of production.

The methodology described in this section was used to determine the STC areas for solar heat integration opportunities identified within FA and FB. Scenarios with STC area to supply the full solar heat demand, that minimised the absolute total difference between solar heat demand and supply, and systems that produced no excess heat were investigated.

Solar fraction

The solar fraction (SF) is a performance indicator commonly used in SHIP studies, and it refers to the fraction of required heat that is provided by the solar heat system. The SF was calculated using the annual heat demand from solar (Q_{demand}) and the total auxiliary heat that was still required from fossil fuels (Q_{aux}), see Equation 4.4.

$$SF = \frac{Q_{demand} - Q_{aux}}{Q_{demand}}$$

Equation 4.4

4.2.4 Preliminary economic analysis methodology

The cost of solar thermal systems in South Africa

Any feasibility study on ST implementation must include an economic study (Kalogirou, 2004). The feasibility of ST technology for process heat depends on the initial cost of the system, the cost of the fuel replaced (Kalogirou, 2003) and the change in fuel cost over the expected life of the project (Joubert et al., 2016). Schnitzer et al. (2007) stated that the profitability is dependent mainly on the investment costs, which in turn are mostly determined by the total STC area.

Joubert et al. (2016) consolidated cost data of large scale (STC area > 10 m^2) ST systems installed in South Africa between 2007 and 2015; data for 47 systems were used to develop a simple model of specific cost as a function of total collector area, see Equation 4.5. SHIP costs are typically expressed in Euro, due to the currency being more stable and to enable direct comparison to the more mature European SHIP industry. The data used included all costs, like components, backup heat sources, commissioning, and maintenance plans for ST installations up to approximately 800 m^2 total collector area. The specific costs of the solar heating system decreased as the size of the system increased, due to economy of scale and the possibility of using larger individual solar collectors (Weiss, 2016).

$$Specific\ system\ cost\ \left[\frac{EUR}{m^2}\right] = -0.41 \times (Collector\ gross\ area\ [m^2]) + 770.34$$

Equation 4.5

The South African SHIP industry is still in its infancy and specific system costs may vary significantly between different applications, and even between different tenders for the same application (Joubert et al., 2016). Economic information for large systems installed in South Africa is limited, with only 6 of the 47 systems included in the model by Joubert et al. (2016) being larger than 400 m^2, and the largest system having a solar collector area only slightly larger than 800 m^2. Due to the linear nature of the cost model (Equation 4.5) and the negative gradient, it makes extrapolation of specific system costs for large installations problematic as negative specific system costs are predicted for systems larger than approximately 1 879 m^2.

However, using the same data set based on South African systems to develop Equation 4.5, Joubert et al. (2016) calculated an average specific system cost of 603 EUR/m^2. Due to the challenges of predicting specific system costs for South African conditions at collector areas larger than 800 m^2, the average cost of 603°EUR/m^2 was rather used to calculate the cost of the solar process heat systems in this study. This value of 603 EUR/m^2 was calculated from the cost data available for 47 large scale installed systems (Joubert et al., 2016).

Net present value

The net present value (NPV) is defined as: the cumulative discounted cash position at the end of the lifetime of a specific project (Turton et al., 2013). The NPV is calculated by taking the sum of the annual cash flows (C_n) discounted to time zero at a specific internal rate (d) over the life of the project (N) see Equation 4.6. The NPV is a good indicator of the feasibility as it represents the total predicted cash flow for the project, discounted to the present value of money, therefore, a positive value means the project is profitable.

$$NPV\ [ZAR] = \sum_{n=0}^{N} \frac{C_n}{(1+d)^n}$$

Equation 4.6

Another parameter used to evaluate the economic viability of a project is the internal rate of return (IRR), which is calculated as the discount rate (d) at which the NPV of an investment is zero (Joubert et al., 2016). If the IRR is greater than the internal discount rate acceptable by a company, then a project is profitable (Turton et al., 2013).

Industry is concerned with the time required to recover the capital cost of an investment. The payback period is a criterion that gives the time required to recover the capital costs with the yearly cash flows discounted to time zero (Turton et al., 2013), which is when the NPV becomes positive.

Levelized cost of heat

The levelized cost of heat (LCOH) is a parameter that provides an estimate of the cost of heat supplied by a heating system over its lifetime, and is commonly used in the description of potential solar process heat systems. The LCOH is only dependent on the cost of the system, therefore, costs savings resulting from the reduced fuel consumption are not included (Joubert et al., 2016). In this study, the LCOH was only calculated for systems that had positive NPV's.

The LCOH is calculated as the ratio of the discounted system costs and the discounted solar heat produced. The economic model used in this study lumped all costs with the capital cost paid at the start, hence, only capital costs were included as system costs, furthermore, only the solar heat utilised by the factory was included. The resulting LCOH equation is shown with Equation 4.7.

$$LCOH\ \left[\frac{ZAR}{kWh}\right] = \frac{Capital\ cost}{\sum_{n=0}^{N} \frac{Q_{demand,n} - Q_{aux,n}}{(1+d)^n}}$$

Equation 4.7

Parameters used in preliminary economic study

A preliminary economic study for each of the identified scenarios were performed, with the parameters used in this study shown in Table 4.1. The system lifetime was taken as 20 years, with annual inflation constant at 6%, which is the upper bound of the South African Reserve Bank's inflation target and therefore presents a maximum estimated inflation rate. The capital costs were assumed to be paid in year zero, and consisted only of the solar process heat system costs, calculated with an average specific cost of 603 EUR/m^2 (Joubert et al., 2016). The Euro to Rand exchange rate was R 14.03 for FA and R 15.20 for FB, during the times when the costs for the different factories were calculated.

In the models, the only annual income was the cost saved due to reduced fuel consumption. The HFO cost per litre was acquired from local suppliers, and that of coal from GEM Commodities World Bank

Group (2017a). The annual cost increases for the fossil fuels were obtained from Joubert et al. (2016). Decommissioning and end-of-life costs were ignored due to insufficient information for the South African SHIP systems.

Table 4.1: Parameters used in preliminary economic study.

Parameter	Value
Inflation (d)	6.00%
System lifetime (N)	20 years
HFO cost per litre	R 7.41
HFO annual cost increase	10.0%
Coal cost per metric ton	R 998.81
Coal annual cost increase	8.80%

4.3 Heat recovery and solar heat in Factory A

4.3.1 Identifying opportunities for heat recovery and solar heat in Factory A

The methodology discussed in section 4.2.1 was applied to FA under base case conditions to identify opportunities for solar heat integration and heat recovery. An illustration of the base case process of FA, showing the stream mass flow rates and temperatures, is shown in Figure 4-4. The illustration in Figure 4-4 aids in the identification of opportunities for heat recovery and solar heat in FA, the description of which follows (stream numbers are stated in brackets):

i.   Heat sinks within the fishmeal production process of FA:

The biggest heat sink within the fishmeal production process of FA was the dryer. With an outlet temperature of 117°C (3 and 4), medium temperature STC's could provide heat to the dryer, however, it was designed to operate with medium (500 kPa gauge) pressure steam at 176°C (A), and producing steam with solar heating systems was not considered in this study. Therefore, the dryer did not present an opportunity for utilisation of solar heat.

The boiler within FA was also a heat sink, however, it required the combustion of fossil fuels and did not present an opportunity for solar heat.

Preheating of fish material was another heat sink. Although preheating did recover some heat from the steam condensate, it was problematic since it necessitated a large volume of condensate within the dryer, which decreased the heat transfer efficiency within the dryer itself and increased the power required by the motor to rotate it. Therefore, FA kept preheating of incoming raw material to a minimum to ensure the dryer operated efficiently, and only diverted 646 kg/h of condensate to preheating, cooling it from 128°C (B) to 67°C (D). Changing the preheating system, or preheating

with solar heat, could improve the efficiency of the dryer and reduce the energy required to dry the materials thereby reducing the steam required from the boiler.

The HFO is maintained at 60°C (F) by an electrical heater, within the temperature range of low temperature STC's. However, heating of HFO was disregarded for this study since the available information was not sufficient to obtain meaningful results.

ii. Streams that could be preheated:

Currently, raw materials enter the process at 8°C (1) and are heated to 45°C (2), which is below 75°C where the cooking process starts. Increasing the preheating temperature or preheating with solar heat could reduce the heating required in the dryer.

iii. Streams that exit the process at high temperature:

The evaporated water exiting the dryer was at approximately 117°C (4), making heat recovery possible. However, FA was in the process of upgrading their seawater condenser and chemical scrubber, therefore, heat recovery from the evaporate was disregarded for this study, as it would influence the operation of the scrubbers, voiding the upgrades.

The fishmeal stream exiting the dryer was also at a high temperature of 117°C (3) which required cooling with seawater, however, heat recovery from the fishmeal stream is problematic since its physical properties are not conducive to heat exchange and the seawater used for cooling was inexpensive, therefore this stream was not deemed feasible for heat recovery.

The steam condensate exiting the dryer (B and C) was at a high temperature. Since this stream was already used for preheating, and decreasing its temperature further could reduce heat losses in the hot well tank, it was considered for heat recovery.

The following opportunities for heat recovery and integrating solar heat in FA were investigated:

i. Drying without preheating
ii. Preheating fish materials with solar heat
iii. Preheating of fish materials with steam condensate

Although drying without preheating does not improve the energy efficiency of FA, it was investigated since the current preheating system used (preheating with steam condensate exiting the dryer, which is then recirculated to the boiler via the hot well tank) conflicts with the efficient operation of the dryer.
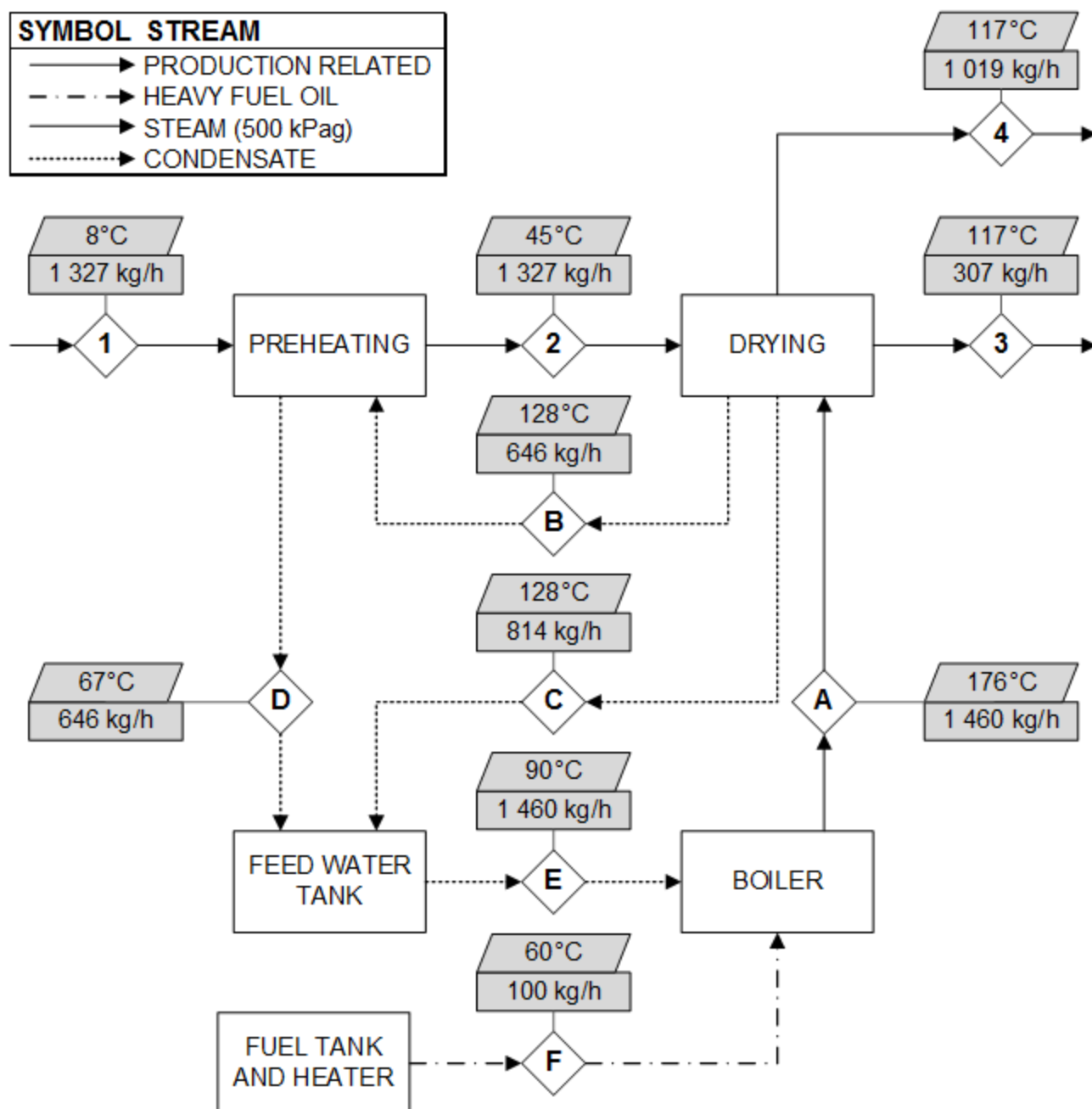
Figure 4-4: The base case process of Factory A, showing stream temperatures and mass flow rates.

4.3.2 Energy and fuel requirements for the different scenarios proposed for Factory A

The effects of the changes listed above on the energy and fuel requirements of the production process at FA were calculated (using the simulations in Microsoft Excel 2016 and Aspen Plus V8.8) and the results are presented in Table 4.2. The base case process values are also shown since all changes were compared to the results obtained for the base case process.

Drying without preheating

In this scenario, raw material entered the dryer at 8°C instead of 45°C and all condensate directly returned to the hot well tank. Compared to the base case process, this scenario required 100 kg/h more steam, thus, even an elementary preheating system could reduce the required steam flow.

Preheating raw material with solar heat

To heat the raw material stream in FA from 8°C to 70°C required 303 MJ/h of energy from the solar process heat system. Cooking of fish materials starts at 75°C (FAO, 1986), during cooking, cell walls rupture, oils are released and proteins are denatured. These occurrences change the physical properties of the fish materials and can make it difficult to handle, especially to transport with equipment such as screw conveyers. Therefore, preheating of the fish materials was only considered to a maximum of 70°C, to ensure that the materials can still be transported easily.

Preheating raw material with steam condensate

Calculations showed that the steam condensate stream contained sufficient energy to heat the raw material stream to 70°C. The energy required from steam was the same as the solar preheating case since the inlet and exit conditions of the dryer were the same. However, the return temperature of the condensate was much lower in this case, resulting in the greater HFO consumption compared to the solar preheating case.

Table 4.2: Energy and fuel requirements for different heat recovery and solar integration scenarios for Factory A.

Scenario	Energy from steam [MJ/h]	Energy from solar [MJ/h]	Steam produced [kg/h]	HFO consumed [litre/h]
Base case process	3 320	0	1 460	100
No preheating	3 547	0	1 560	103
Preheating with solar heat	3 169	303	1 394	92
Preheating with steam condensate	3 169	0	1 394	99

4.3.3 Total collector area for solar opportunities in Factory A

Solar thermal collector technology to be used

The solar process heat system had to supply water at 85°C for the opportunities identified in FA, since the highest temperature to which materials had to be heated was 75°C and a temperature difference of 10°C between the inlet temperature of the heating water and exit temperature of the cold stream was deemed adequate for sufficient heat transfer rates. FPC's are well suited to this temperature range (high solar irradiation conversion efficiencies are achievable), although ETC's are also well

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suited. FPC's were selected to be investigated for FA, due to their low cost and simplicity of design and operation.

Available solar irradiation and roof area at Factory A

Values of the monthly solar irradiation on FA's location was obtained from TMY results generated in Meteonorm 7, since two options for installing the collector were possible, the selection was based on the incident irradiation amounts. The annual horizontal irradiation at FA's location was 1 982 kWh/m^2. The solar heat system must be installed on the roof of FA, which has a pitch of 30°. The roof has North-East (46° East of North) and South-West (46° West of South) facing planes, both with an approximate area of 3 400 m^2. The annual irradiation on the North-East plane was 2 130 kWh/m^2, and 1 525 kWh/m^2 on the South-West plane. Due to the greater annual solar irradiation, the North-East facing roof area was selected for calculating the required solar collector area. The TMY results are shown in Appendix D.

Total solar collector area for Factory A

The total STC area required for preheating the raw material stream in FA was significantly less for systems with SF < 1, compared to the system that fully supplied the solar heat demand (SF = 1). The required STC areas for the proposed systems, calculated as shown in section 4.2.3, are shown in Table 4.3. To supply the entire solar heat demand required a system with a total collector area of 782 m^2, however, SF's of 0.81 and 0.73 were achieved with systems having collector areas of 49% (384 m^2) and 43% (337 m^2) of the full supply system area, respectively. The SF of the proposed systems, at 0.81 and 0.73, was similar to the values of 0.6 to 0.8 that typically results in economically viable systems for South African factories with a constant heat demand throughout the year (Hess, 2016a), although this is not the only factor affecting economic viability. Of the proposed systems in Table 4.3, option A results in the minimum difference between solar heat demand and supply, while option B produces no excess heat. Additional results of the solar heat supply calculations as well as the energy requirements of the altered base case process (for options with SF < 1) are shown in Appendix D.

Table 4.3: Total solar thermal collector area of the solar thermal heat systems proposed for preheating the raw material stream in Factory A.

Proposed system	Total solar thermal collector area [m^2]	Calculated solar fraction
Full supply	782	1.00
Option A	384	0.81
Option B	337	0.73

Output of the solar process heat system for raw material preheating in Factory A

Due to the solar irradiation on FA reaching a minimum in June, and the solar heat demand remaining fairly constant through the year, all systems that attempt to have a very high solar fraction (≈ 1) will be grossly oversized outside of the winter months and generate large amounts of unused heat throughout the year. The solar heat demand for preheating the raw material stream at FA, as well as the heat supplied by the proposed solar heat systems are shown in Figure 4-5. The system that supplied the full solar heat demand, delivered excess heat during all months except June, where it matched the solar heat demand. The solar irradiation profile shown is for a North-East facing plane with a 30° slope.

The proposed solar systems, options A and B, better matched the solar heat demand of FA. As can be seen in Figure 4-5, the heat supplied by option A was slightly higher than the demand at the beginning and end of the year, and less than the demand in the middle of the year. The heat supplied by this system was expected to be greater than the demand during some months and less than during others, since its area was selected to minimise the difference between demand and supply. Option B, which produced no excess solar heat, matched the demand during January and was less than the monthly demand for the remainder of the year, where the solar irradiation was lower.
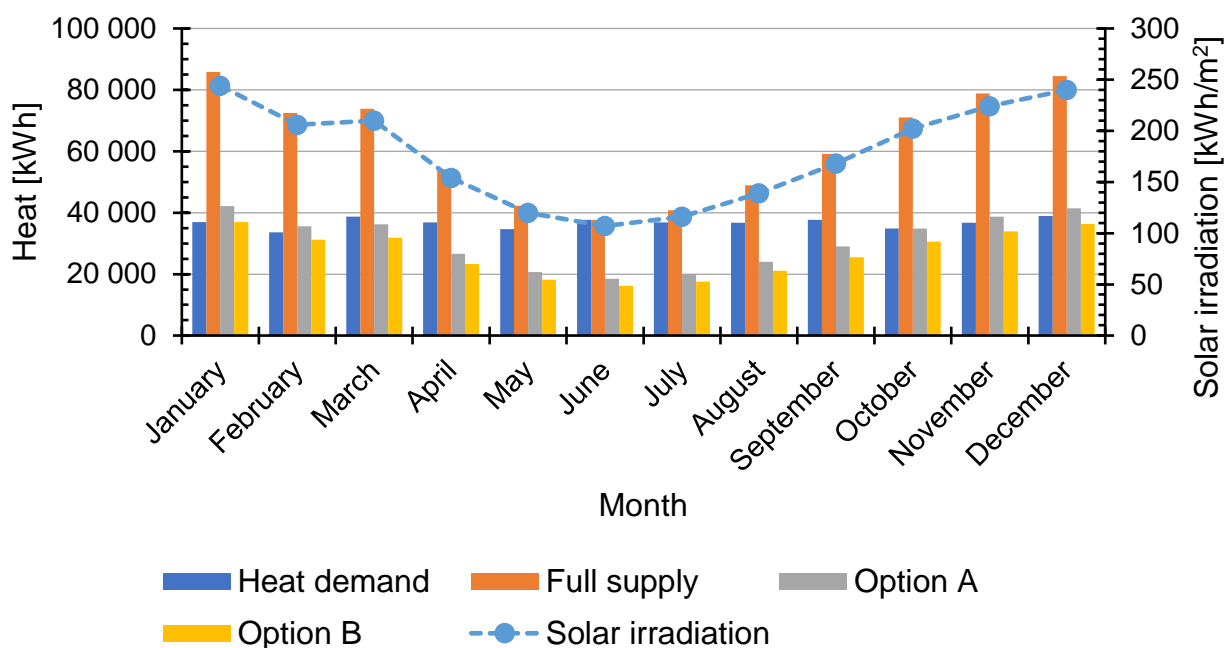
Figure 4-5: Solar heat demand for preheating the raw material steam in Factory A, solar heat supplied by various solar heat systems and, the solar irradiation profile at Factory A's location for a plane facing 46° East of North at a slope of 30°. The total collector areas were: 782 m² for the full supply, 384 m² for option A and 337 m² for option B.

4.4 Heat recovery and solar heat in Factory B

4.4.1 Identifying opportunities for heat recovery and solar heat in Factory B

The methodology discussed in section 4.2.1 was applied to FB under base case conditions to identify opportunities for solar heat integration and heat recovery. An illustration of the base case process of FB, showing the stream mass flow rates and temperatures, is shown in Figure 4-6. The illustration in Figure 4-6 aids in the identification of opportunities for heat recovery and solar heat in FB, the description of which follows (stream numbers are stated in brackets):

i. Heat sinks within the fishmeal and fish oil production process of FB:

The cookers and dryers were the biggest heat sinks in the fishmeal and fish oil production process of FB. The average outlet temperature of the cooking stage was 99°C (2), while the outlet temperatures of the dryers were 90°C (11) and 97°C (14), for the first and second drying stages respectively. Medium temperature STC's could provide heat at the required temperatures for these units, however, the units were designed to utilise medium temperature steam, and producing steam with solar heating systems was not considered in this study. Therefore, the cookers and dryers did not present an opportunity for utilisation of solar process heat.

The boilers that provide steam for FB are also heat sinks, these require the combustion of fossil fuels, therefore, they did not present an opportunity for solar process heat.

ii. Streams that could be preheated:

Raw materials enter the process at 22°C (1), which was significantly lower than the operating temperature of the cookers. Low temperature STC's could be used to preheat this stream, this could significantly reduce the energy required by the cookers.

Stickwater concentrate at 50°C (10 and 13) was mixed with solids at a calculated 83°C (6), and the combined stream exited the first drying stage at 90°C (11). The stickwater temperature was relatively low and heating this stream could reduce the energy required during drying.

The water that replenishes the condensate lost in the steam production system entered the process at 20°C (A), and preheating this stream would reduce the energy required by the boiler to produce steam.

iii. Streams that exit the process at high temperature:

The fish oil product stream has the potential for heat recovery (16). The temperature of this stream is not measured by FB since it does not affect the process, therefore, the exact temperature is unknown. However, the temperature is believed to be within the range of 65°C (8) to 83°C (6),

based on available upstream stream temperatures. Heat recovery from the fish oil product was investigated, however, it is unlikely that this stream would be used to heat other streams, since it contains the final saleable fish oil product and there is a risk of contamination. Additionally, the mass flow rate of this stream at 1 509 kg/h is low compared to that of the other streams in this process, which limits the amount of transferable heat.

Fishmeal exits the final drying stage at 97°C (14), which is quite high. The fishmeal is in the form of solids and recovering heat, or transferring heat to another stream would be difficult to do efficiently, therefore, heat recovery from the fishmeal product was disregarded in this study.

Neither the stickwater (8), nor the evaporate from the dryers (9 and 12) were considered in this study, since heat was already recovered in the multi-stage evaporators and interference could be detrimental to its operation.

The following opportunities for heat recovery and integrating solar heat in FB were investigated:

- i. Preheating fish materials before cooking
- ii. Reheating stickwater concentrate prior to mixing with the press cake and grax stream
- iii. Preheating the make-up water to the hot well tank
- iv. Heat recovery from the fish oil product stream
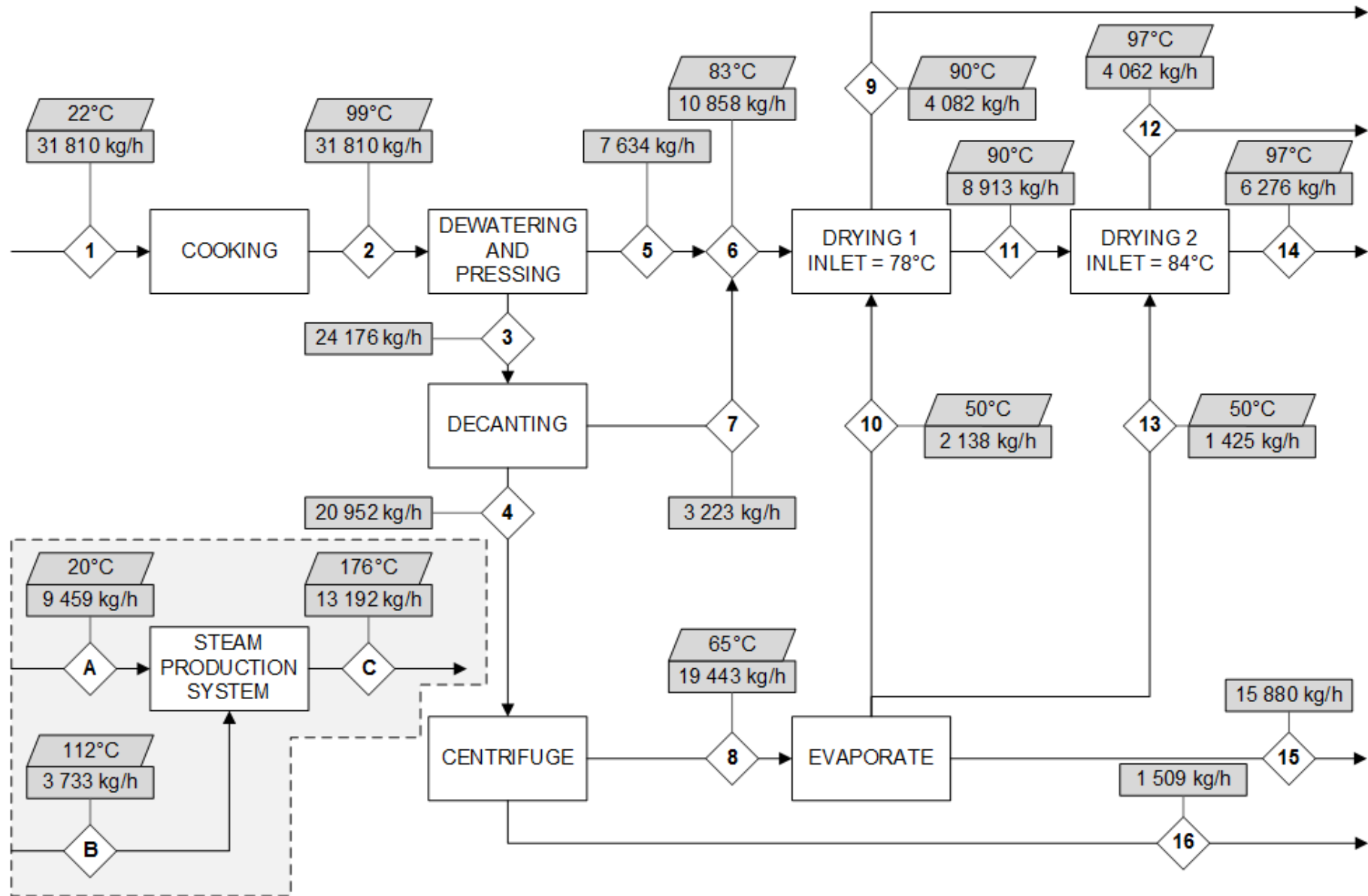
Figure 4-6: The base case process of Factory B, showing stream temperatures and mass flow rates. Stream A is the make-up water stream, stream B is recovered condensate at 50 kPa (gauge) and stream C is live steam at 800 kPa (gauge).

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4.4.2 Energy and fuel requirements for the different scenarios proposed for Factory B

The effects of the changes listed above on the energy and fuel requirements of the production process at FB were calculated (using the simulations in Microsoft Excel 2016 and Aspen Plus V8.8) and the results are presented in Table 4.4. The base case process values are also shown since all changes were compared to the results obtained for the base case process. Additional results for the individual sections in the production process and the steam production system are shown in Appendix E.

Preheating of fish before cooking

Heating the raw material from 22°C to 70°C required 5 547 MJ/h from the solar process heat system, and significantly reduced the energy required by the cookers. Preheating of fish material only to temperature of 70°C was investigated due to the complications in material handling once materials start cooking at 75°C. Compared to the base case process, the energy required from live steam was reduced by 19.6% and the coal consumption reduced by 18.7%.

Reheating stickwater concentrate before drying

Reheating the stickwater concentrate ensured a higher inlet temperature to the dryers and less energy required during drying. Reheating the concentrate to 75°C was investigated, since this is achievable by low temperature collectors and minimises the possibility of thermal degradation of the proteins in the concentrate. With the concentrate at 75.0°C the inlet temperatures to the first and second drying stages were calculated as 81.8°C and 87.6°C respectively, which is slightly higher than the base case process temperatures of 77.5°C and 83.5°C.

Heating the stickwater concentrate required 306 MJ/h of solar heat and only slightly reduced the steam flow and fuel consumption. Compared to the base case process the total mass flow rate of steam was reduced by 1.1% and the coal consumption also by 1.1%. The energy required from live steam was reduced by only 1.1%, thus, the overall effect on the energy requirements is significantly less compared to the fish preheating case since only a small amount of the materials is heated by a few degrees.

Preheating make-up water to the steam production system

Preheating the boiler make-up water to 75°C from 20°C was investigated and it was found that this required 2 197 MJ/h of heat from the solar heating system. Preheating to 75°C is desirable since it is within the operating range of low temperature collectors and FB stated that higher temperatures would result in cavitation within the boiler feed pumps.

Preheating the boiler make-up water does not involve the fishmeal production process and therefore the mass flow rate of steam remained constant compared to the base case process. However, the

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higher inlet temperature to the boilers resulted in less energy required to produce steam, and compared to the base case process the coal consumption was reduced by 6.4%.

Combined fish and make-up water preheating

It is possible to use solar heat to preheat both the raw material and boiler make-up water streams, consequently, the combination of these two scenarios were investigated and found to result in the largest reduction in coal consumption. Preheating both streams required 7 332 MJ/h of solar heat and compared to the base case process the coal consumption was reduced by 23.8%. The total coal savings are not the sum of the savings of the individual scenarios, since raw material preheating reduces the required steam mass flow rate, which reduces the amount of make-up water required, consequently, the make-up water preheating is also affected.

Table 4.4: Energy and fuel requirements for different solar integration scenarios for Factory B.

Scenario	Energy from steam [MJ/h]	Energy from solar [MJ/h]	Steam produced [kg/h]	Coal consumed [kg/h]
Base case	28 316	0	13 192	2 078
Fish preheating	22 769	5 547	10 733	1 689
Stickwater concentrate reheating	28 010	306	13 047	2 054
Make-up water preheating	28 316	2 197	13 192	1 945
Make-up water and fish preheating	22 769	7 332	10 733	1 582

Heat recovery from the fish oil product stream

The only stream in FB that presented an opportunity for heat recovery, without adversely affecting other units in the process, was the fish oil product stream. The temperature of the fish oil stream exiting the oil polisher is not measured and therefore was assumed to be within the range of 65°C to 83°C. The lower limit is the temperature at which the stickwater entered the evaporators, and the upper limit is the temperature of the press cake and grax before it was mixed with concentrate and entered the first drying stage. The excess heat in this stream can be recovered by preheating the raw material stream entering the cookers or the stickwater concentrate prior to drying. Calculations for heat recovery from fish oil was based on the higher temperature as this was the most favourable scenario for heat recovery.

The recoverable heat from the fish oil stream represents a small proportion of the heat required for preheating by FB. A minimum temperature difference of at least 10°C was ensured by setting the final temperature of the fish oil stream to 30°C when preheating the raw material, and 60°C when

heating the stickwater concentrate. For the fish oil flowing at 1 509 kg/h, 164 MJ/h of heat is released when it is cooled to 30°C from 83°C, this is approximately 3% of the heat required to preheat the raw material stream, resulting in a temperature increase of 1°C. Cooling the fish oil to 60°C requires removing 72 MJ/h of heat, this is approximately 23% of the heat required to heat the stickwater concentrate, resulting in a temperature increase of 6°C.

Heat recovery from the fish oil stream was therefore not deemed feasible as even under the most favourable conditions, only a very small proportion of the overall process heat requirement can be obtained via heat recovery from this stream. Further, the actual temperature of this stream was estimated as 83°C, however, it could be lower than this value, resulting in less recoverable heat. In the event of equipment failure, a valuable product would be contaminated and would most likely have to be discarded, which is undesired. Therefore, heat recovery from this stream was not investigated further.

4.4.3 Total collector area for solar opportunities in Factory B

Solar thermal collector technology to be used

FPC's and ETC's would both be well suited to heat materials to the required temperatures of 70°C and 75°C. Due to their low cost and simplicity FPC's were selected to be investigated for FB.

Solar irradiation and available area at Factory B

The installation of the STC's at FB was not limited to a certain size or orientation since the facility has multiple roof areas and is surrounded by large open spaces. For ST installations a good preliminary value is obtained by having collectors face the equator, thus, due North in the Southern hemisphere and due South in the Northern hemisphere (Duffie and Beckman, 2013). A good initial value for the collector slope is to have it approximately equal to the latitude of the location, solar collectors are installed at an incline to maximise the amount of incident beam radiation (Souka and Safwat, 1966). Therefore, a plane facing North with an inclination of 35° was investigated.

The solar irradiation values at the location of FB were obtained from the TMY results generated in Meteonorm 7. Annual horizontal solar irradiation at FB's location was 2 025 kWh/m^2. The annual solar irradiation on the plane facing due North with an inclination of 35° is 2 258 kWh/m^2, this value was used to calculate the STC area for FB. The TMY results are shown in Appendix E.

Solar collector area for Factory B

Systems capable of supplying the entire solar heat demand of the identified applications in FB, would be very large. The calculated total STC areas of the proposed systems are shown in Table 4.5. The STC areas calculated for separate fish preheating and make-up water preheating systems were

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12 220 m^2 and 4 841 m^2, respectively, and 16 153 m^2 for a combined system. Even for the global ST industry, these would be large systems, considering that the three largest SHIP plants installed in the world has collector areas ranging from 7 804 m^2 to 39 300 m^2 (Weiss et al., 2017).

The calculated total STC areas were significantly less for systems with SF < 1, compared to systems that fully supplied the solar heat demand (SF = 1) for the different applications. Of the proposed systems shown in Table 4.5, option A results in the minimum difference between solar heat demand and supply, while option B produces no excess heat during months of production (i.e. January, November and December were excluded). Option A had a SF of 0.38, while the area was reduced by 86% compared to the full supply system, for all applications. Option B had a SF of 0.12, while the area was reduced by 96% compared to the full supply system. For the South African industry these would be relatively large systems since very few systems installed in South Africa are larger than 400 m^2 (Joubert et al., 2016).

The considerable reductions in total STC area, could place options A and B in a more favourable position from an economic viability perspective than the full supply system. Since the collectors comprise most of the capital cost of a ST heat system (Schnitzer et al., 2007), the capital cost of options A and B would be significantly less than the full supply system.

Table 4.5: Total solar thermal collector areas and solar fractions of the proposed solar heat systems for the different applications in Factory B. Option A was calculated to minimise the difference between solar heat demand and supply, and option B produces no excess heat during months of production.

Proposed system:	Full supply	Option A	Option B
Solar fraction:	1.00	0.38	0.12
Application	**Area [m^2]**	**Area [m^2]**	**Area [m^2]**
Fish preheating	12 220	1 751	503
Stickwater concentrate reheating	673	96	28
Make-up water preheating	4 841	694	199
Make-up water and fish preheating	16 153	2 314	665

Output of the solar process heat system for raw material preheating in Factory B

Preheating of the raw material stream in FB with solar heat, was the single scenario that resulted in the greatest reduction in coal consumption (Table 4.4), therefore, it was selected as an example to show the solar heat demand and supply. The solar heat demand for preheating the raw material stream at FB, as well as the heat supplied by the proposed solar heat systems, and the available solar irradiation are shown in Figure 4-7. The solar irradiation values shown is for FB's location on a plane facing North with a slope of 35°.

The amount of excess heat produced by the full supply system is exacerbated by the fact that the solar heat demand reaches a maximum in June, when the available solar irradiation is at its lowest. The

full supply system produces no excess heat during June, while producing exorbitant amounts of unused heat during the remainder of the year, especially January, November and December when no FMFO production takes place, due to no fish being landed at the factory.

Due to the very high heat demand during April to June compared to the rest of the year, solar heat systems that do not attempt to match the demand during this period are significantly smaller and produce less excess heat during other times of the year. The proposed solar heat systems, options A and B, produce significantly less heat than the full supply system, however, the amount of excess unused heat produced is also significantly less. The heat supplied by option A was less than the demand for most of the year, except February and October, since it resulted in the difference between demand and supply being minimised. The heat supplied by option B matched the demand during October, and was less than the demand during the rest of the year, when fishmeal production occurred.
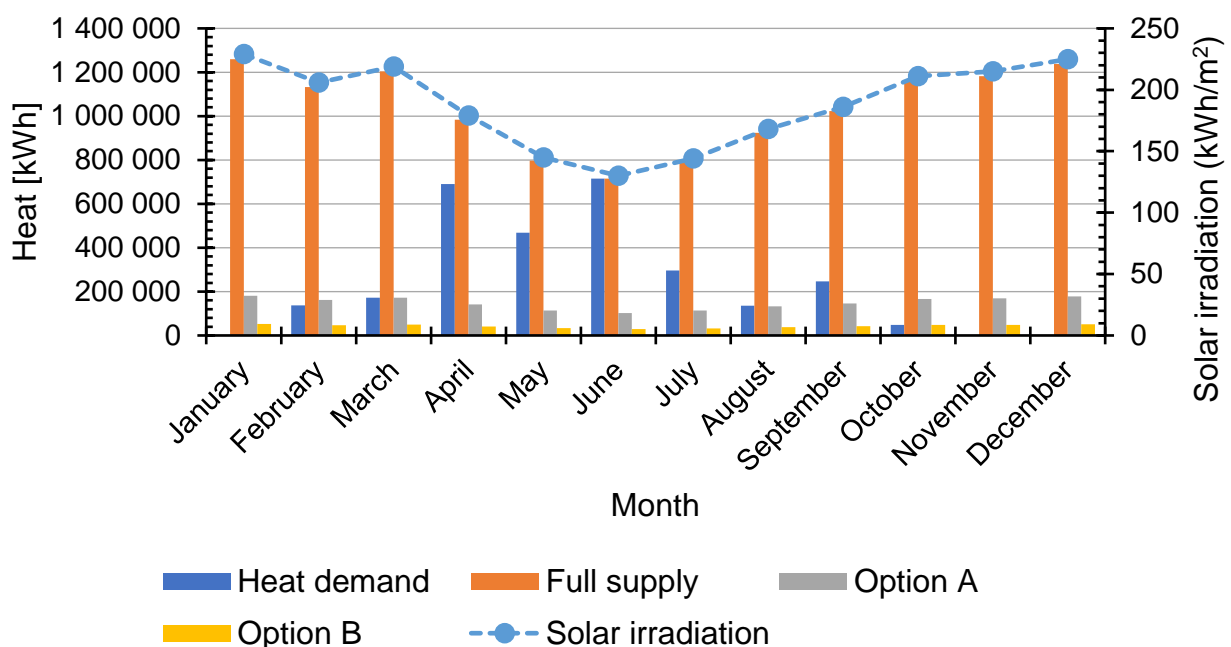
Figure 4-7: Solar heat demand for preheating the raw material steam in Factory B, solar heat supplied by various solar heat systems and the solar irradiation profile at Factory B's location for a plane facing North with a slope of 35°. The total collector areas were: 12 220 m² for the full supply, 1 751 m² for option A and 503 m² for option B.

4.5 Preliminary economic analysis

The viabilities of the proposed solar heat integration and heat recovery options were quantified by performing a preliminary economic study. The NPV was used as the main indicator of economic viability, with a positive value indicating a profitable option. Although financial calculations dominate industrial decisions, decision makers should consider the additional benefits of integrating ST heat: investment in renewable energy will improve the reputation and public profile of the

company, and it will improve energy security (Schnitzer et al., 2007). However, these aspects are typically not addressed from an engineering perspective and are therefore not discussed in this study.

4.5.1 Factory A economic analysis

<u>Economic viability of heat recovery and solar heat scenarios</u>

All the heat integration options proposed for FA were economically viable, however, the solar heat systems were the most attractive options from an economical perspective. The results of the preliminary economic analysis for FA are shown in Table 4.6. All of the proposed systems had positive NPV's, which ranged from R 0.2 million for heat recovery from condensate to R 3.3 million for the solar system with an area of 384 m^2, therefore, all three scenarios were economically viable. Despite the large amounts of excess heat produced, and the large capital cost of R 6.6 million, the system that supplied the full heat demand was also economically viable with an NPV of R 2.1 million, this was due to the large cost savings achieved resulting from a significant reduction in HFO consumption. The most promising scenario was the solar preheating system: option A, as it had the highest NPV and IRR at 3.3 million Rand and 13.2% respectively.

Although the proposed solar heat systems, options A and B, had lower HFO savings compared to the full supply system (see Table 4.6), they had larger NPV's. This was due to the significantly reduced capital costs, at 48% and 42% of that of the full supply system, resulting in lower expenses at the start of the project. The detailed NPV calculation of option A is shown in Appendix F.

The capital cost of the condensate heat recovery system included two pumps, insulated piping, modifications to the current steam sleeves and the necessary control equipment. A first estimate of R 1.1 million was obtained from the cost of the pumps predicted by Aspen Economic Analyser (Aspen Plus V8.8). The detailed NPV calculation of the heat recovery system is shown in Appendix F.

The payback periods for all the proposals were long, ranging from 11.8 to 17.4 years, which is problematic as industry typically prefers payback periods of less than five years. However, it is known that the payback periods for solar process heat systems are long, due to the high capital costs (Joubert et al., 2016).

The LCOH of option A and B for solar preheating was significantly less than that of the system supplying the full demand, however, these were still greater than the LCOH of the condensate heat recovery system. The LCOH values calculated for the various proposed systems are also shown in Table 4.6. The LCOH of the solar options A and B was less than the full supply system as a result of a greater proportion of the heat produced being utilised by FA. The lowest LCOH obtained for a solar heat system was that of option B at R 0.77, which was slightly less than the R 0.79 of option A

since it did not produce any unused heat. The lowest LCOH achieved was that of the condensate heat recovery system at R 0.22, which was considerably lower than the other systems since it produced the same amount of utilised heat than the full supply solar heat system, while having a significantly lower capital cost.

Table 4.6: Results of preliminary economic study on proposed heat recovery and solar heat integration systems for Factory A.

Scenario	Capital cost [R millions]	Net present value [R millions]	Internal rate of return	Payback period [years]	Levelized cost of heat	Volume HFO saved [Litres]
Full supply (782 m^2)	6.6	2.1	8.6%	16.3	R 1.31	42 994
Option A (384 m^2)	3.2	3.3	13.2%	11.8	R 0.79	32 061
Option B (337 m^2)	2.8	2.8	13.0%	11.9	R 0.77	27 790
Condensate heat recovery	1.1	0.2	7.8%	17.4	R 0.22	6 694

The investigated opportunities for solar heat integration and heat recovery identified in FA were mutually exclusive, thus, only solar preheating or heat recovery from the condensate can be implemented as either one will eliminate the possibility of the other. However, while a combination of solar heating and heat recovery could also occur in reality, only the limits of the two options (i.e. only solar heat or only heat recovery) were investigated in this study. Heat recovery from the condensate stream increased the energy efficiency of the plant, however, it will not result in the greatest fuel savings and was not the most promising investment.

Sensitivity of Factory A economic analysis findings

Since the solar heat systems were the most promising from an economic perspective, and the costs of these systems in the South African market has yet to be established (Joubert et al., 2016), a sensitivity analysis was performed as part of the preliminary economic analysis of FA. The sensitivity of the NPV and IRR, of the solar preheating systems of FA, to changes in the specific system cost and fuel price increase was investigated. The specific system cost and fuel price increase were varied between -50% and +50% of the original value (Table 4.1), since this resulted in a reasonable range being covered for both parameters.

The NPV and IRR of the proposed solar heat systems decreased as the specific cost increased, due to greater capital costs. The results from varying the specific cost of the solar process heat system is shown in Figure 4-8. The profitability of all the solar process heat systems proposed for FA increased

at the specific cost of 301.5 EUR/m² (scenario with -50% of average specific cost), with the NPV of options A and B increasing by approximately 50% compared to the values calculated for the current average of 603 EUR/m². At a specific system cost of 904.5 EUR/m² (scenario with +50% of average specific cost), the full supply solar heat system was no longer viable, with a negative NPV of R 1.2 million, while the NPV's of both options A and B decreased by approximately 50%. Over the specific cost range investigated, options A and B remained economically viable throughout, with IRR values remaining greater than the specified inflation of 6%.

The specific cost range of 301.5 EUR/m² to 904.5 EUR/m² investigated, which was obtained by varying the original specific cost of 603 EUR/m² by 50% and 150% of its value is realistic for the South African market. Multiple systems installed in South Africa achieved a specific cost of 301.5 EUR/m², while several other installed systems had a specific cost even greater than 904.5 EUR/m² (Joubert et al., 2016). Thus, a sufficient range of possible system costs were investigated.
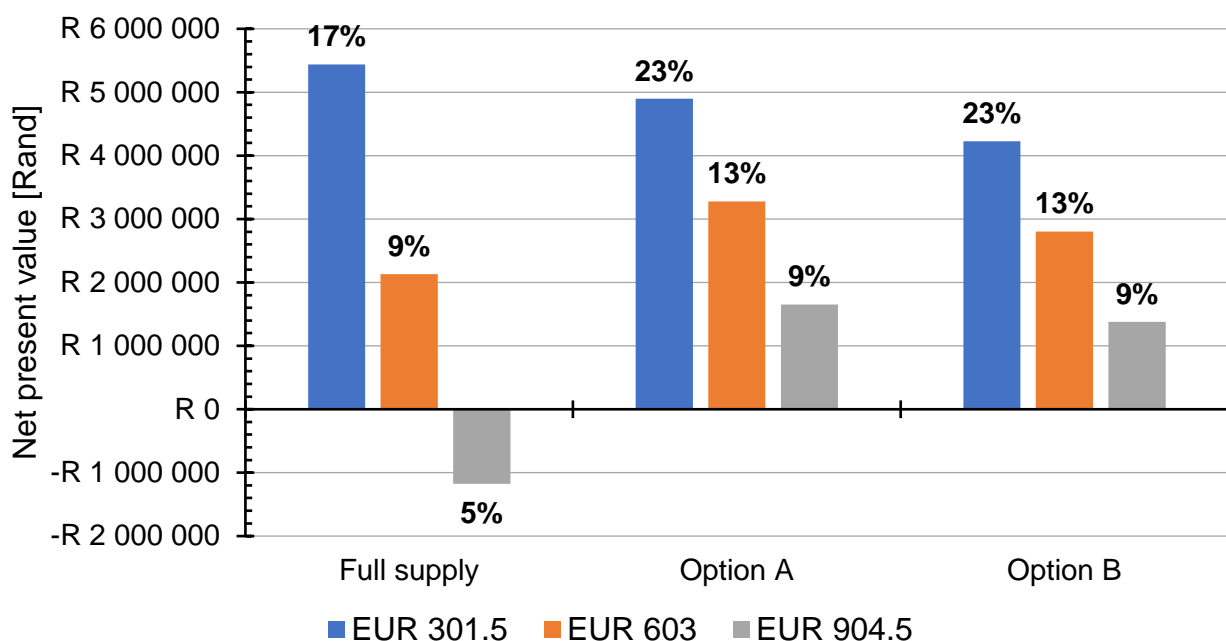
Figure 4-8: Net present value and internal rate of return of Factory A solar preheating systems for 50%, 100% and 150% of the original 603 EUR/m² specific system cost. Values shown above net present value columns are the internal rates of return. The total collector areas were: 782 m² for the full supply, 384 m² for option A and 337 m² for option B.

The NPV and IRR increased as the annual fuel-price-increase increased, due to the greater projected cost savings. The results from varying the annual fuel price increase is shown in Figure 4-9. At a fuel price increase of 5% annually (scenario with -50% of original annual fuel increase), the full supply solar heat system was no longer economically viable, with a negative NPV of R 1.1 million. However, both options A and B remained economically viable throughout the range of fuel price increase investigated, with IRR values remaining greater than the specified inflation of 6%.

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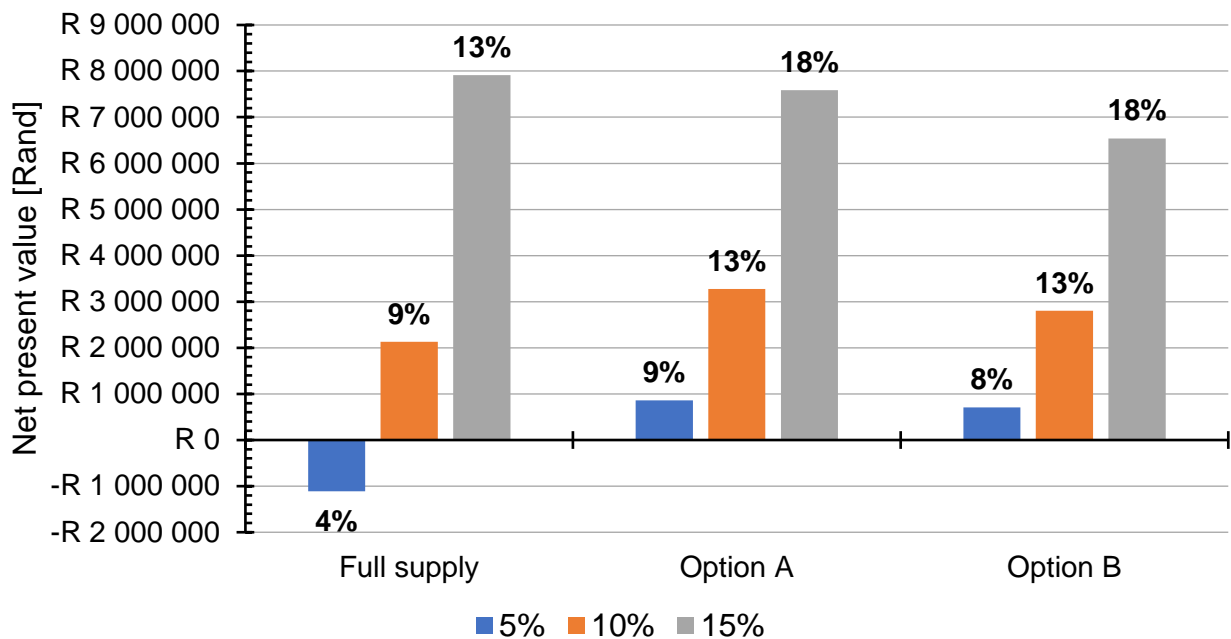
Figure 4-9: Net present value and internal rate of return of Factory A solar preheating systems for 50%, 100% and 150% of the original annual fuel price increase. Values shown above net present value columns are the internal rates of return. The total collector areas were: 782 m² for the full supply, 384 m² for option A and 337 m² for option B.

A qualitative comparison of the results show that the IRR was more sensitive to the specific system costs, while the NPV was more sensitive to the annual fuel price increase. Of the two factors investigated, varying the specific system cost resulted in the largest IRR values, while varying the annual fuel price increase resulted in the largest NPV values.

4.5.2 Factory B economic analysis

Economic viability of solar heat integration in Factory B

Neither the proposed solar heat systems nor the full supply system for FB were economically viable. The results of the preliminary economic analysis for solar heat integration in FB are shown in Table 4.7. The NPV's for all the applications were negative, regardless of the solar collector area, thus none were economically viable. The NPV's of both options A and B were significantly less negative than that of the full supply system. The low cost of the fuel (coal) used by FB is suggested to be one of the main factors affecting the viability of the systems. Since the annual cost savings arising from reduced fuel consumption are not sufficient to offset the large capital costs incurred, over the lifetime of the system. Detailed NPV calculations for the option B systems in all applications are shown in Appendix F.

Table 4.7: Net present values and amounts of coal saved for the proposed solar heat systems of the different applications in Factory B.

Application	Full supply		Option A		Option B	
	NPV [R millions]	Coal saved [tons]	NPV [R millions]	Coal saved [tons]	NPV [R millions]	Coal saved [tons]
Fish preheating	-94.1	733	-9.2	280	-2.4	92
Stickwater concentrate reheating	-5.1	45	-0.4	18	-0.1	7
Make-up water preheating	-38.3	250	-4.0	96	-1.0	32
Make-up water and fish preheating	-125.2	935	-12.5	357	-3.2	117

The fish preheating and boiler make-up water preheating by solar heat as two separate, exclusive scenarios were further investigated. It is anticipated that the combined system where both heating duties are performed simultaneously with solar heating would be too vulnerable to fluctuations within the process, due to fluctuating raw material amounts, therefore, it was not further investigated. The 'option B' systems, where no excess heat is produced during production months, were selected to be investigated as these had the most potential of becoming economically viable (least negative NPV's).

Coal price and specific system costs to ensure economically viable systems

Solar heat systems to preheat fish and boiler make-up water in FB would be viable if the cost of coal were R 2 053 and R 2 320 per metric ton, respectively. This has not occurred in the last 20 years, and assuming an average annual price increase of 8.8%, coal prices will reach these values in the next 8 to 10 years.

A system specific cost of 293 EUR/m^2 was required for the fish preheating system, and 260 EUR/m^2 for the make-up water preheating, for these systems to become economically viable. Similar specific costs have been reported for systems installed in South Africa (Joubert et al., 2016), however, it is highly unlikely that these specific costs will be achievable for a large industrial heat installation that will most likely require stainless steel piping and equipment. FB requires stainless steel piping and equipment due to the installation's proximity to the sea, which promotes corrosion of equipment and piping.

Effect of heat transfer inefficiencies in the process units

The heat transfer efficiency of cookers and dryers in the base case process was defined as 100% due to a lack of information. This represents the minimum required energy, with inefficiencies increasing

the amount of energy required from steam. The system sizing and economic analysis were re-performed with equipment efficiencies of 80% and 60%, and the 'option B' ST systems. These additional analyses were performed for the fish and make-up water preheating systems, since they had the largest individual solar heat demands, the results are shown in Table 4.8.

Even with equipment efficiencies as low as 60% the fish preheating with solar heat system was not economically viable, as indicated by the negative NPV's. As the heat transfer efficiency of cooking and drying equipment decreased the economic viability of the fish preheating solar heat system increased. Since the effect of reducing the energy required from steam in the cookers, by supplying a portion with ST heat via preheating, became more pronounced. Changing the efficiency did not change the production process, therefore, the energy required from solar heat and the STC area remained constant.

As the equipment efficiencies decreased the energy required from steam increased, this required an increase in the amount of steam produced, consequently, the solar process heat system for boiler make-up water preheating needed to be resized. The solar collector areas required to preheat make-up water to the boiler, with equipment (cooker and dryer) efficiencies of 80% and 60% were 249 m^2 and 332 m^2 respectively. The increased collector area resulted in higher capital costs, therefore, the NPV of the boiler water preheating system decreased as the equipment efficiency decreased.

Table 4.8: Net present value for fish and make-up water preheating with solar heat for Factory B, based on the solar system that produced no excess heat (option B). Efficiency values of 60% and 80% refer to the efficiency of transferring energy from live steam to the material inside the process units.

Scenario	NPV for efficiency of 80% [R millions]	NPV for efficiency of 60% [R millions]
Fish preheating	-1.8	-0.9
Make-up water preheating	-1.3	-1.7

The above results show that, if the heat transfer efficiencies of the cooker and dryer were incorporated into the base case process, that it would not have affected the economic viability of the proposed solar process heat systems. Since FB used equipment specifically made to produce fishmeal it is believed that the equipment efficiency would be relatively high and closer to 80% like the dryer in FA, than 60%.

The effect of fuel type and solar fraction on economic viability of the fish preheating system

The very low cost of coal, compared to other fossil fuels, was one of the main reasons the solar heat systems proposed for FB were not economically viable. The base case process was based on coal since it is the cheapest fuel and FB had sufficient capacity from coal fired boilers to produce the required steam. However, FB also had a HFO-fired boiler, the effect of using HFO instead of coal,

at the same boiler-burner efficiency as in reality was investigated. In order to do this, the amount of coal saved was converted to an energy amount and the volume of HFO required to produce an equal amount of energy (by combustion) was then calculated and used as the amount of fuel saved.

For most SF's, the fish preheating solar heat system would be economically viable, if FB used HFO as main fuel source. The NPV of fish preheating systems replacing heat from coal and HFO at various solar fractions are shown in Figure 4-10. The system replacing heat produced from coal is not economically viable for any of the SF's investigated and would appear to not be viable at all. The system replacing heat produced from HFO is economically viable up to a SF of 0.9, and appears to reach a maximum NPV between SF's of 0.5 and 0.75. This system is economically viable instead of the scenario where coal is the main fuel source, due to the considerably higher cost of energy produced by HFO per MJ, compared to coal. Detailed results for this section is shown in Appendix F.

As the SF increases the amount of unused heat also increases, eventually reaching a point where the system cost of producing the additional heat outweighs the cost savings of the fossil fuel saved. To demonstrate this, the fraction unused heat of the total amount of solar heat produced is also shown in Figure 4-10. At a SF of 0.05, 30% of the generated solar heat is wasted, while at a solar fraction of 1, 77% of the generated heat is wasted, which is a significant amount. Since the heat produced from coal was cheaper than that from solar over all SF's, the NPV became increasingly negative as the solar fraction increased. The cost of heat produced from HFO was more expensive than that from solar and thus the NPV increased with increasing SF, until it reached a point where the capital cost of the system, to ensure the additional capacity, outweighed the income from fuel savings.
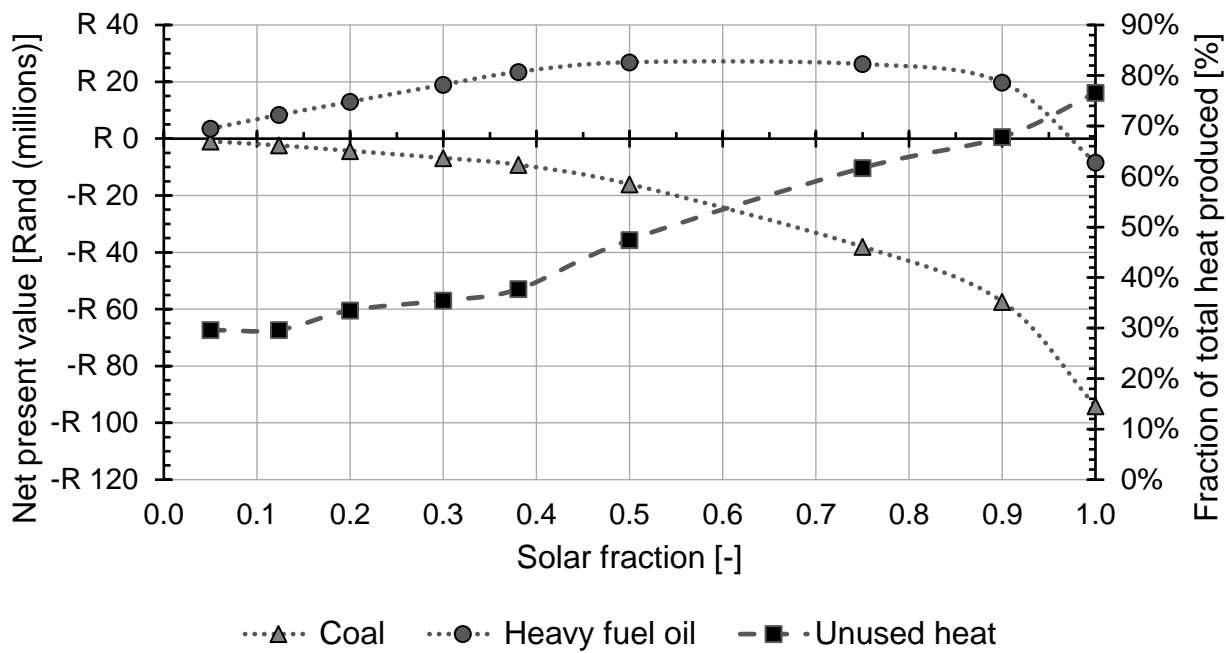
Figure 4-10: Net present value for the fish preheating solar heat systems with coal and heavy fuel oil as main fuel sources, plotted against the solar fraction. The fraction unused heat of the total solar heat produced is also shown on the right-hand axis.

5 <u>Conclusions and recommendations</u>

This chapter concludes this study. The success of the study in reaching its aim is established and conclusions are drawn from the results. Recommendations with regards to the integration of solar process heat into fishmeal factories in general, and specifically to the two factories studied, are made. Recommendations for additional work to compliment this study are given.

5.1 Status of this study

The lack of information about the energy consumption and potential for solar process heat in South African fishmeal factories motivated the aim of this study, which was to determine the feasibility of integrating solar heat into the production processes of South African fishmeal factories. The success of this study in achieving this aim is discussed in terms of the objectives that were set out:

  i.   Data representative of production at two South African fishmeal factories were collected and used to developed base case processes specific to each factory. The factories studied used distinctly different production methods and the scale of production between the factories was significantly different.

 ii.   Production at FA used a single dryer, which was understandable since it is a small factory. almost exclusively processing lean fish, however, this is not a common fishmeal production method used internationally. The equipment and operating conditions of FB was typical of the wet-pressing fishmeal production method commonly used internationally.

iii.   Opportunities for preheating raw materials with solar heat and heat recovery from steam were identified and quantified for FA. Opportunities for preheating raw materials, reheating stickwater concentrate and preheating make-up water to the boiler by solar heat were identified and quantified for FB.

 iv.   Flat plate collectors were identified as a suitable ST collector technology since water at approximately 85°C was required. A method to optimise the STC area by minimising the difference between solar heat demand and supply was developed and used to determine the total STC area required in the different scenarios, and a STC area that results in no excess heat produced during months of production was also investigated.

  v.   A preliminary economic analysis was performed and the results indicated that all the options proposed for FA were economically viable, while none of the options proposed for FB were economically viable.

All the objectives have been achieved, therefore, this study is seen as a successful first investigation into the feasibility of integrating solar process heat into existing production processes in South Africa.

Fishmeal factories were investigated as an example industry due to their typically high energy demands.

5.2 Conclusions

The technical feasibility of a solar process heat system, to supply a significant portion of the heat demand, increases as the process heat demand is similar in profile to the available solar irradiation. The annual production schedule of FA was relatively constant, which resulted in a solar fraction of 0.81 for a solar process heat system with an area that minimised the difference between solar heat demand and supply, and 0.73 for a system that produced no excess heat. In contrast, production at FB reached a maximum during the winter months when the solar irradiation was at its lowest, the result was a solar fraction of 0.38 for a system with minimised difference between solar heat demand and supply, and 0.12 for a system that produced no excess heat in any of the FMFO production months. The fairly constant solar heat demand of FA was more similar to the profile of solar irradiation than that of FB, consequently, higher SF's were achieved for FA.

The cost of the fuel used to provide the process heat for each specific factory was a major factor to the viability of solar process heat systems. All the options proposed for FA were economically viable since solar heat reduced the amount of relatively expensive HFO consumed. Although significant coal savings could be achieved by incorporating solar process heat into FB, these proposals were not economically viable due to the combination of high capital costs of installing solar heating and the low cost of coal. However, for the scenario where FB used HFO as main fuel source instead of coal, the proposed systems were economically viable over almost the entire SF range investigated.

The fishmeal factories studied were energy intensive, furthermore, it was difficult to quantify the losses in these factories accurately due to a lack of information. Due to relatively high profit margins on FMFO, energy consumption and expenditure on fuel are not presently major concerns for the respective factories. In future, in scenarios where the cost of fuel increases and financial penalties related to the use of fossil fuels may be implemented, factories are expected to start addressing inefficiencies in their processes to reduce fuel consumption.

Production at FB was with the wet-pressing method used internationally, therefore, the integration options proposed for FB could be valid for a large amount of fishmeal factories globally. ST technology can replace a fraction of the energy required in fishmeal factories with solar process heat and would thereby reduce the fuel consumption of these factories.

5.3 Recommendations to fishmeal factories

5.3.1 General recommendations

The fluctuating production rates and uncertainty regarding future production amounts at the factories make planning a solar process heat system challenging. However, it was observed that there were other hot water requirements on the same premises of the factories, e.g. for washing of equipment or staff ablutions, which tend to be more constant and predictable. These were outside of the scope of this study, however, a solar process heat system installed at either factory could be used to provide heat for these purposes which could aid in increasing the economic viability of the systems. It is suggested that the factories investigate the possibility of alternative uses of warm water from solar heat systems (e.g. to provide warm water for general use), as this should be less costly than using steam condensate or water heated by electricity.

5.3.2 Recommendations to Factory A

The incorporation of solar process heat at FA was the most promising in this study. If a constant production schedule throughout the year could be ensured, which is equivalent to that of the base case process, then this preliminary study, in terms of efficiency and cost, indicates that a solar process heat system of 384 m^2 used for preheating the raw material stream, would be a profitable investment, with a predicted NPV of 3.3 million Rand and IRR of 13.2%. This system provides 81% of the preheating heat demand and the payback period is predicted as 11.8 years.

5.3.3 Recommendations to Factory B

The main fuel source used, and annual production schedule were the biggest factors affecting the economic viability of solar heat systems at FB. Due to the low cost of coal, none of the solar heat options proposed for FB were viable, however, the scenarios were viable when HFO were used. If FB increases the use of the HFO fired boiler in future, this study should be revisited, as a solar process heat system may then be economically viable. The fact that there is no production and consequently no solar heat demand during January, November and December, when solar irradiation is at its highest, results in significant amounts of heat produced that is not utilised. This was a major factor for the proposed systems for FB not being economically viable.

The recovery of steam condensate by FB, at 28.3% was low, and resulted in the stream fed to the boiler being at a low temperature due to large amounts of make-up water being required. Greater condensate recovery would result in a higher temperature stream entering the boiler and less fuel

consumed to produce steam. Therefore, FB could investigate methods to decrease steam losses from the system during normal operation.

5.4 Recommendations for further study

Based on the results of this study the following opportunities for further study have been identified:

- Identify the effect of modifying the production process of FA to include a cooking and water separation stage, with or without further water evaporation, on the product quality and energy consumption of the factory. The current process used by FA uses a single dryer to evaporate all the water that needs to be removed, while a large proportion of this water could be separated with mechanical means which consume significantly less energy. As these alternative methods will split-up the material flow and produce streams with different protein content, the additional cost of including evaporators to concentrate stickwater and recover the protein in this stream must be justified in terms of the increased fishmeal yield.

- Use the solar heat demand and production schedules determined in this study, and simulate the solar heat system in appropriate software like T*SOL, to obtain more accurate values of the solar process heat system efficiency and output.

- Little detail on the design of the solar heat systems was given in this study, this was accounted for by using an average cost calculated for a variety of system concepts and designs. A more accurate costing should be obtained by performing a detailed design of the most promising systems. Standard costs for equipment have not been established for the South African market, consequently, suppliers will have to be directly contacted to obtain the costs for these systems.

These recommendations would be a valuable addition to the work done in this study, and would aid the specific factories in obtaining an implementable solution.

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Appendices

i

Appendix A: Industries suitable to solar thermal heat

Although certain types of processes occur in nearly all industries there are some that have a greater potential for SHIP than others (Schnitzer et al., 2007). The practicality of implementing SHIP depends not only on the operating temperature but also on the specific operation. It is much easier to heat a fluid stream by implementing a counter current heat exchanger for example than to heat solid objects like bottles or metal cylinders. There are various industries where high temperature processes dominate, such as the steel and ceramic industries. In these industries there is a surplus of heat and heat for lower temperature processes is obtained from heat recovery and renewable technologies will not be considered (Schnitzer et al., 2007).

There are various industries that consume heat at low to medium temperatures that are ideally suited to ST technologies (Kalogirou, 2003). Appendix Table 1 is a summary of industrial processes in various industries that utilise process heat at temperature levels that are suited to ST heat. The food industry is particularly promising for the implementation of SHIP with various processes occurring below 100°C and large amounts of energy being consumed for sterilisation, pasteurisation and cleaning operations annually.

Appendix Table 1: Indicative temperature ranges of industrial processes suited to ST implementation. Information from Kalogirou (2003).

Industry	Process	Temperature [°C]
Dairy	Pressurisation	60 – 80
	Sterilisation	100 – 120
	Drying	120 – 180
	Concentrates	60 – 80
	Boiler feed water	60 – 90
Tinned food	Sterilisation	110 – 120
	Pasteurisation	60 – 80
	Cooking	60 – 90
	Bleaching	60 – 90
Textile	Bleaching, dyeing	60 – 90
	Drying, degreasing	100 – 130
	Dyeing	70 – 90
	Fixing	160 – 180
	Pressing	80 – 100
Paper	Cooking, drying	60 – 80
	Boiler feed water	60 – 90
	Bleaching	130 – 150

Industry	Process	Temperature [°C]
Chemical	Soaps	200 – 260
	Synthetic rubber	150 – 200
	Processing heat	120 – 180
	Pre-heating water	60 – 90
Meat	Washing, sterilisation	60 – 90
	Cooking	90 – 100
Beverages	Washing, sterilisation	60 – 80
	Pasteurisation	60 – 70
Flours and by-products	Sterilisation	60 – 80
Timber by-products	Thermodifussion beams	80 – 100
	Drying	60 – 100
	Pre-heating water	60 – 90
	Preparation pulp	120 – 170
Bricks and blocks	Curing	60 – 140
Plastics	Preparation	120 – 140
	Distillation	140 – 150
	Separation	200 – 220
	Extension	140 – 160
	Drying	180 – 200
	Blending	120 – 140

Appendix B: Supplement to Factory A base case process

Appendix Table 2: Weekly report data collected from Factory A from September 2015 to September 2016.

Date	Product [ton]	Raw material [ton]	Moisture	Fuel consumed [litre]	Water consumed [m³]
24 September 2016	34	145	4.9%	11 422	220
17 September 2016	20	110	4.1%	6 640	180
10 September 2016	27	110	4.1%	8 978	180
03 September 2016	17	74	3.6%	6 750	170
27 August 2016	24	110	4.1%	8 904	195
06 August 2016	46	180	3.9%	19 333	305
30 July 2016	32	126	4.9%	10 682	220
23 July 2016	29	129	4.2%	10 334	205
16 July 2016	24	110	3.8%	9 002	215
09 July 2016	32	119	3.8%	11 466	205
30 June 2016	34	164	4.9%	11 016	210
25 June 2016	34	130	4.2%	10 743	205
18 June 2016	20	93	3.8%	6 771	289
11 June 2016	25	118	4.9%	8 308	190
07 June 2016	62	285	4.9%	20 777	381
28 May 2016	34	131	3.7%	10 924	205
21 May 2016	44	203	4.8%	15 900	324
14 May 2016	34	140	3.9%	11 750	190
07 May 2016	38	147	4.1%	12 604	190
30 April 2016	24	118	4.8%	8 326	160
23 April 2016	34	136	4.2%	10 820	160
16 April 2016	29	127	3.9%	9 850	160
09 April 2016	26	135	4.9%	9 200	170
02 April 2016	19	73	4.1%	5 952	185
26 March 2016	26	112	3.9%	8 912	265
19 March 2016	20	96	4.9%	6 078	195
12 March 2016	29	129	4.8%	8 598	202
05 March 2016	28	127	4.2%	9 264	195
27 February 2016	27	133	4.9%	8 945	190
20 February 2016	25	114	4.2%	7 038	181
13 February 2016	28	121	3.9%	8 203	173

Date	Product [ton]	Raw material [ton]	Moisture	Fuel consumed [litre]	Water consumed [m³]
06 February 2016	39	157	4.1%	10 114	173
30 January 2016	43	195	4.5%	11 021	342
09 January 2016	19	70	3.9%	5 883	190
02 January 2016	15	76	4.8%	5 369	90
19 December 2015	43	178	4.1%	16 138	215
05 December 2015	46	210	4.4%	15 852	276
28 November 2015	29	124	4.1%	9 020	201
21 November 2015	28	106	4.9%	8 495	190
14 November 2015	30	137	4.1%	10 131	209
07 November 2015	23	93	3.9%	6 828	305
31 October 2015	34	161	4.2%	9 217	285
17 October 2015	51	194	3.5%	14 916	372
10 October 2015	30	133	4.8%	8 834	212
03 October 2015	37	152	3.8%	11 424	169
26 September 2015	21	105	4.5%	5 715	116
19 September 2015	28	99	3.2%	8 413	216

Appendix Table 3: Process control sheet data collected from Factory A from 20 September to 23 September 2016. Only the cooker outlet temperatures are shown, since during this time the steam pressure remained constant at 500 kPa (gauge), condensate pressure remained constant at 150 kPa (gauge) and the outlet temperature of the hot well tank remained constant at 90°C.

Date	Time	Temp [°C]	Date	Time	Temp [°C]	Date	Time	Temp [°C]
2016/09/20	14:00	112.1	2016/09/21	20:30	114.8	2016/09/22	22:00	117.2
2016/09/20	14:30	123.4	2016/09/21	21:00	115.3	2016/09/22	22:30	119.0
2016/09/20	15:00	125.1	2016/09/21	21:30	112.9	2016/09/22	23:00	119.4
2016/09/20	15:30	121.1	2016/09/21	22:00	114.1	2016/09/22	23:30	118.8
2016/09/20	16:00	122.1	2016/09/21	22:30	113.8	2016/09/23	00:00	115.1
2016/09/20	16:30	121.1	2016/09/21	23:00	116.5	2016/09/23	00:30	113.0
2016/09/20	17:00	119.9	2016/09/21	23:30	119.3	2016/09/23	01:00	116.9
2016/09/20	17:30	116.7	2016/09/22	00:00	117.0	2016/09/23	01:30	113.8
2016/09/20	18:00	118.0	2016/09/22	00:30	114.5	2016/09/23	02:00	112.5
2016/09/20	18:30	115.0	2016/09/22	01:00	112.2	2016/09/23	02:30	111.9
2016/09/20	19:00	114.7	2016/09/22	01:30	111.0	2016/09/23	03:00	114.4
2016/09/20	19:30	117.5	2016/09/22	02:00	113.7	2016/09/23	03:30	116.8

Date	Time	Temp [°C]	Date	Time	Temp [°C]	Date	Time	Temp [°C]
2016/09/20	20:00	118.8	2016/09/22	02:30	114.8	2016/09/23	04:00	117.7
2016/09/20	20:30	115.4	2016/09/22	03:00	117.1	2016/09/23	04:30	115.1
2016/09/20	21:00	116.5	2016/09/22	03:30	115.0	2016/09/23	05:00	114.3
2016/09/20	21:30	118.1	2016/09/22	04:00	116.9	2016/09/23	05:30	116.0
2016/09/20	22:00	117.9	2016/09/22	04:30	118.1	2016/09/23	06:00	118.1
2016/09/20	22:30	115.5	2016/09/22	05:00	117.7	2016/09/23	06:30	116.1
2016/09/20	23:00	117.0	2016/09/22	05:30	116.4	2016/09/23	07:00	115.4
2016/09/20	23:30	112.5	2016/09/22	06:00	115.6	2016/09/23	07:30	116.7
2016/09/21	00:00	114.5	2016/09/22	06:30	115.0	2016/09/23	08:00	113.6
2016/09/21	00:30	113.7	2016/09/22	07:00	114.7	2016/09/23	08:30	119.4
2016/09/21	01:00	112.0	2016/09/22	07:30	114.2	2016/09/23	09:00	117.7
2016/09/21	01:30	109.2	2016/09/22	08:00	118.0	2016/09/23	09:30	118.6
2016/09/21	02:00	108.1	2016/09/22	08:30	119.0	2016/09/23	10:00	119.4
2016/09/21	02:30	110.8	2016/09/22	09:00	118.1	2016/09/23	10:30	118.1
2016/09/21	03:00	110.0	2016/09/22	09:30	119.3	2016/09/23	11:00	119.4
2016/09/21	03:30	112.5	2016/09/22	10:00	120.9	2016/09/23	11:30	118.7
2016/09/21	04:00	114.0	2016/09/22	10:30	123.2	2016/09/23	12:00	116.9
2016/09/21	04:30	116.2	2016/09/22	11:00	121.8	2016/09/23	12:30	117.4
2016/09/21	05:00	114.1	2016/09/22	11:30	121.0	2016/09/23	13:00	117.9
2016/09/21	05:30	112.3	2016/09/22	12:00	118.9	2016/09/23	13:30	118.5
2016/09/21	06:00	109.0	2016/09/22	12:30	117.4	2016/09/23	14:00	117.6
2016/09/21	09:00	114.9	2016/09/22	13:00	118.7	2016/09/23	14:30	118.7
2016/09/21	09:30	117.8	2016/09/22	13:30	118.1	2016/09/23	15:00	117.4
2016/09/21	10:00	119.0	2016/09/22	14:00	117.9	2016/09/23	15:30	118.6
2016/09/21	10:30	116.9	2016/09/22	14:30	118.6	2016/09/23	16:00	119.4
2016/09/21	11:00	115.3	2016/09/22	15:00	117.1	2016/09/23	16:30	118.9
2016/09/21	11:30	116.9	2016/09/22	15:30	112.1	2016/09/23	17:00	117.4
2016/09/21	12:00	117.0	2016/09/22	16:00	113.4	2016/09/23	17:30	120.3
2016/09/21	12:30	116.1	2016/09/22	16:30	115.0	2016/09/23	18:00	119.8
2016/09/21	13:00	114.6	2016/09/22	17:00	117.1	2016/09/23	18:30	117.2
2016/09/21	13:30	119.1	2016/09/22	17:30	114.9	2016/09/23	19:00	114.1
2016/09/21	14:00	120.1	2016/09/22	18:00	116.6	2016/09/23	19:30	116.7
2016/09/21	14:30	120.9	2016/09/22	18:30	115.1	2016/09/23	20:00	116.9
2016/09/21	15:00	121.8	2016/09/22	19:00	117.7	2016/09/23	20:30	115.7

Date	Time	Temp [°C]	Date	Time	Temp [°C]	Date	Time	Temp [°C]
2016/09/21	15:30	123.9	2016/09/22	19:30	119.5	2016/09/23	21:00	114.4
2016/09/21	16:00	124.6	2016/09/22	20:00	118.9	2016/09/23	21:30	113.1
2016/09/21	19:00	116.9	2016/09/22	20:30	117.1	2016/09/23	22:00	115.9
2016/09/21	19:30	116.1	2016/09/22	21:00	119.5	2016/09/23	22:30	113.5
2016/09/21	20:00	118.0	2016/09/22	21:30	118.5			

Preheating

IN	OUT
T_{in} 8°C	T_{out} 45°C
M_{in} 1 327 kg/h	M_{out} 1 327 kg/h
COMPOSITION IN:	**COMPOSITION OUT:**
Protein 15.5%	Protein 15.5%
Water 78.0%	Water 78.0%
Fat 2.3%	Fat 2.3%
Ash 4.2%	Ash 4.2%

Heat required per component:

Component:	Q [MJ/h]	M [kg/h]	$Cp\Delta T$ [MJ/kg]
Protein	15.48	205.23	0.08
Water	158.05	1034.80	0.15
Fat	2.32	31.03	0.07
Ash	2.34	55.61	0.04
TOTAL	**178.20 MJ/h**		

Appendix Figure 1: Simulation of the preheating section of the base case process of Factory A in Microsoft Excel 2016.

Drying		Evaporate
IN	**OUT**	117°C
		1 019 kg/h

T_{in}	45°C	T_{out}	117°C
M_{in}	1 327 kg/h	M_{out}	307 kg/h

COMPOSITION IN:		COMPOSITION OUT:	
Protein	15.5%	Protein	66.8%
Water	78.0%	Water	5.0%
Fat	2.3%	Fat	10.1%
Ash	4.2%	Ash	18.1%
			100.0%

Heat required per component:

Component:	Q [MJ/h]	M [kg/h]	CpΔT [MJ/kg]
Protein	30.93	205.23	0.15
Fat	4.62	31.03	0.15
Ash	4.88	55.61	0.09

Water heating and evaporation:

Q [MJ/h]	2618.75

TOTAL:	**2659.17 MJ/h**

Appendix Figure 2: Simulation of the drying section of the base case process of Factory A in Microsoft Excel 2016.

96501221 CR 10-14 50 Hz
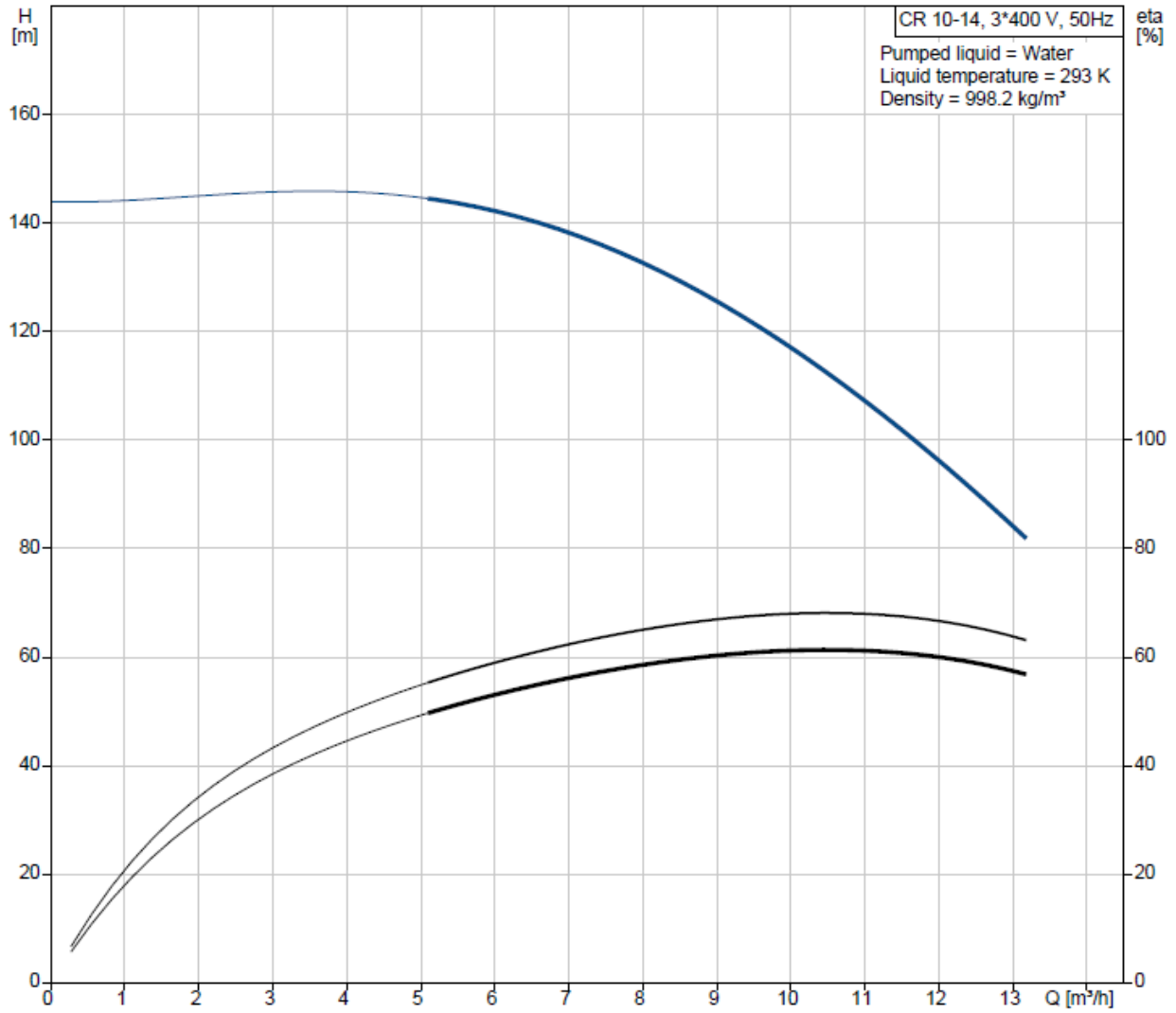
Appendix Figure 3: Pump curve of the boiler feed pump (Grundfos A96501221) used in Factory A.

Appendix C: Supplement to Factory B base case process

Appendix Table 4: Daily production summary data collected from Factory B for the 2016 production period.

Date	Production time [hour]	Raw material processed [tons]	Fishmeal produced [tons]	Fish oil produced [kg]
09 February 2016	5.00	49.30	12.00	1 343
11 February 2016	43.33	368.28	76.00	8 122
14 February 2016	3.75	30.00	18.00	1 247
15 February 2016	10.75	101.79	22.74	3 100
21 February 2016	60.75	514.53	120.92	8 054
22 February 2016	20.25	437.81	115.53	8 099
24 February 2016	38.00	349.89	99.00	7 402
25 February 2016	16.00	192.22	52.59	2 348
27 February 2016	5.00	41.51	16.70	3 055
28 February 2016	6.00	80.00	23.00	
03 March 2016	12.00	129.31	27.00	1 382
06 March 2016	71.00	333.10	67.00	
09 March 2016	67.50	365.32	94.00	5 333
10 March 2016	22.50	334.03	112.00	695
13 March 2016	42.00	677.87	156.00	4 712
14 March 2016	7.50	85.00	34.00	
17 March 2016	14.00	215.13	39.00	2 726
21 March 2016	78.50	884.54	239.00	19 775
22 March 2016	6.00	104.49	11.00	753

x

Date	Production time [hour]	Raw material processed [tons]	Fishmeal produced [tons]	Fish oil produced [kg]
23 March 2016	8.00	52.45	20.30	1 823
03 April 2016	71.50	1 157.59	284.00	14 665
06 April 2016	8.00	215.17	26.00	4 156
07 April 2016	21.50	654.03	153.16	20 029
10 April 2016	69.50	1 562.22	344.00	45 238
13 April 2016	79.00	1 052.30	281.00	37 985
17 April 2016	75.00	2 672.14	573.00	76 935
18 April 2016	30.00	1 004.72	218.00	15 156
20 April 2016	48.00	1 142.85	339.80	58 989
21 April 2016	24.00	628.01	195.00	3 600
24 April 2016	72.00	602.70	155.00	32 174
25 April 2016	24.00	871.15	191.00	18 507
27 April 2016	47.00	1 194.46	229.00	12 266
28 April 2016	24.00	673.33	140.00	8 910
02 May 2016	96.00	1 088.27	270.37	29 253
03 May 2016	23.00	888.45	186.00	21 897
04 May 2016	24.00	436.37	73.00	
05 May 2016	24.00	903.54	163.00	17 915
08 May 2016	72.00	1 137.48	286.00	22 959
09 May 2016	20.00	250.59	77.00	3 730
11 May 2016	39.42	275.58	64.00	
15 May 2016	60.00	628.90	145.00	14 561
16 May 2016	27.00	661.71	160.00	
17 May 2016	15.50	311.23	77.00	

Date	Production time [hour]	Raw material processed [tons]	Fishmeal produced [tons]	Fish oil produced [kg]
18 May 2016	17.25	339.28	83.00	
23 May 2016	76.00	1 523.12	349.00	22 670
06 June 2016	21.00	189.79	66.00	7 686
07 June 2016	24.00	817.91	170.00	13 873
08 June 2016	24.00	802.88	222.00	12 623
09 June 2016	21.50	714.32	139.80	11 150
12 June 2016	72.00	1 344.94	180.00	15 114
13 June 2016	24.00	451.68	129.00	3 602
14 June 2016	24.00	615.36	156.50	7 977
16 June 2016	64.00	1 003.64	216.00	12 330
19 June 2016	72.00	2 012.43	469.00	24 628
21 June 2016	32.00	788.54	178.58	16 201
23 June 2016	48.00	1 192.60	273.50	14 322
26 June 2016	74.00	2 508.02	491.43	28 408
29 June 2016	15.00	391.76	89.00	6 942
30 June 2016	21.50	813.62	175.00	12 456
05 July 2016	91.00	704.19	164.00	
14 July 2016	58.00	691.49	147.50	4 724
18 July 2016	96.00	856.44	225.50	40 000
20 July 2016	52.00	847.52	147.00	
24 July 2016	64.00	1 059.01	266.00	
27 July 2016	51.00	1 552.11	331.00	22 175
11 August 2016	63.50	991.10	255.00	3 648
17 August 2016	9.17	22.27	5.20	
21 August 2016	16.50	416.21	96.70	2 521
01 September 2016	69.50	1 183.13	278.00	11 708
09 September 2016	67.50	1 314.82	318.00	20 237
15 September 2016	75.75	1 631.74	384.00	

Date	Production time [hour]	Raw material processed [tons]	Fishmeal produced [tons]	Fish oil produced [kg]
04 October 2016	40.00	312.49	84.00	617
06 October 2016	33.50	609.56	127.00	
04 November 2016	54.75	23.42	5.50	
12 November 2016	19.75	10.76	2.50	

Appendix Table 5: Steam production data collected from Factory B. Only data entries where coal only was used are shown, since these values were used to determine the coal consumption in the base case process of Factory B.

Date	Coal consumed [Tons]	Water [m³]	Condensate recovered [%]	Condensate recovered [m³]	Steam produced [Tons]
08 February 2016	19	11	49	62	113.4
09 February 2016	9	60	22	15	61.2
10 February 2016	26	78	60	109	163.8
12 February 2016	10	77	0	0	70.2
14 February 2016	19	138	0	0	123.3
15 February 2016	19	106	10	13	122.4
20 February 2016	35	320	15	52	315.9
24 February 2016	30	38	0	1	191.7
25 February 2016	24	30	23	53	211.5
26 February 2016	15	114	0	0	96.3
01 March 2016	22	180	0	0	142.2
02 March 2016	31	199	13	28	198.9
03 March 2016	33	165	29	67	210.6
07 March 2016	29	231	13	31	222.3
08 March 2016	21	152	0	0	136.8
09 March 2016	35	220	11	27	223.2
15 March 2016	18	17	19	24	113.4
16 March 2016	10	23	17	12	63.9
17 March 2016	36	156	38	96	226.8
18 March 2016	17	72	39	47	107.1
19 March 2016	20	61	57	80	126.9
20 March 2016	55	262	32	122	345.6

Date	Coal consumed [Tons]	Water [m³]	Condensate recovered [%]	Condensate recovered [m³]	Steam produced [Tons]
21 March 2016	43	142	57	173	270.9
23 March 2016	34	292	7	20	261.0
05 April 2016	25	28	0	0	150.3
12 April 2016	40	23	46	118	232.2
13 April 2016	40	23	54	167	276.3
14 April 2016	43	30	56	227	368.1
02 May 2016	70	337	52	341	593.1
31 May 2016	19	166	13	24	168.3
06 July 2016	30	168	5	7	135.9
08 July 2016	20	172	26	55	189.9
09 July 2016	30	124	42	60	127.8
10 July 2016	17	64	47	58	110.7
11 July 2016	33	148	2	3	135.0
12 July 2016	24	153	6	9	145.8
13 July 2016	34	171	29	71	218.7
14 July 2016	54	256	56	298	481.5
18 July 2016	125	494	45	392	788.4
19 July 2016	60	222	48	199	372.6
20 July 2016	50	130	24	42	154.8
21 July 2016	50	181	1	1	160.2
22 July 2016	60	183	54	197	330.3
23 July 2016	60	104	62	174	251.1
25 July 2016	75	247	54	275	461.7
26 July 2016	80	206	58	281	438.3
27 July 2016	40	231	2	4	229.5
05 August 2016	29	305	46	228	448.2
06 August 2016	38	287	49	268	491.4
09 August 2016	25	246	19	55	255.6
10 August 2016	50	171	55	210	342.9
11 August 2016	45	262	26	83	286.2
21 August 2016	26	89	55	106	171.9
31 August 2016	30	155	42	120	260.1
19 September 2016	30	32	3	6	205.2
20 September 2016	25	52	0	0	86.4

Date	Coal consumed [Tons]	Water [m³]	Condensate recovered [%]	Condensate recovered [m³]	Steam produced [Tons]
21 September 2016	25	93	0	0	78.3
22 September 2016	100	164	47	140	269.1
23 September 2016	100	70	46	60	117.9
24 September 2016	60	65	34	34	90.0
03 October 2016	40	87	5	4	78.3
04 October 2016	91	135	40	198	444.6
10 October 2016	20	131	0	0	117.9
11 October 2016	21	139	0	0	119.7
18 October 2016	18	126	0	0	99.9
02 November 2016	15	30	35	11	27.9
09 November 2016	20	127	0	0	103.5
10 November 2016	20	73	0	0	76.5
16 November 2016	30	123	0	0	99.0
17 November 2016	10	35	0	0	30.6
18 November 2016	40	84	0	0	75.6
21 November 2016	13	108	11	13	103.5
23 November 2016	15	107	0	0	89.1
24 November 2016	9	96	0	0	83.7
25 November 2016	8	196	0	0	175.5

Raw material information:

Type	Anchovy
User rate	1 000 kg/h

Anchovy composition:

Dry matter	16%
Oil	7%
Water	77%
Protein	13%
Ash	3%

IN → **COOKING** → **OUT**

Raw material flow options:

Use typical values	⦿
Use own values	○
Flow rate	31 810 kg/h

22°C	**Temperature**	98.5°C

Energy required to cook materials:

$$\dot{m} \qquad \int_{T_1}^{T_2} Cp\, dT$$

	$\dot{m}$	$\int_{T_1}^{T_2} Cp\, dT$	Heat req.	Heat req.
Protein	4 171 kg/h	159 kJ/kg	6.62E+05 kJ/h	662 MJ/h
Ash	919 kg/h	91 kJ/kg	8.37E+04 kJ/h	84 MJ/h
Oil	2 227 kg/h	157 kJ/kg	3.50E+05 kJ/h	350 MJ/h
Water	24 494 kg/h	317 kJ/kg	7.77E+06 kJ/h	7769 MJ/h
			8.86E+06 kJ/h	8864.3 MJ/h

Estimate of steam required:

Gauge pressure	180 kPa (g)
Absolute pressure	281 kPa
Energy transferred	2 172 kJ/kg
Steam flow	4 081 kg/h

Appendix Figure 4: Simulation of the cooking section of the base case process of Factory B in Microsoft Excel 2016.
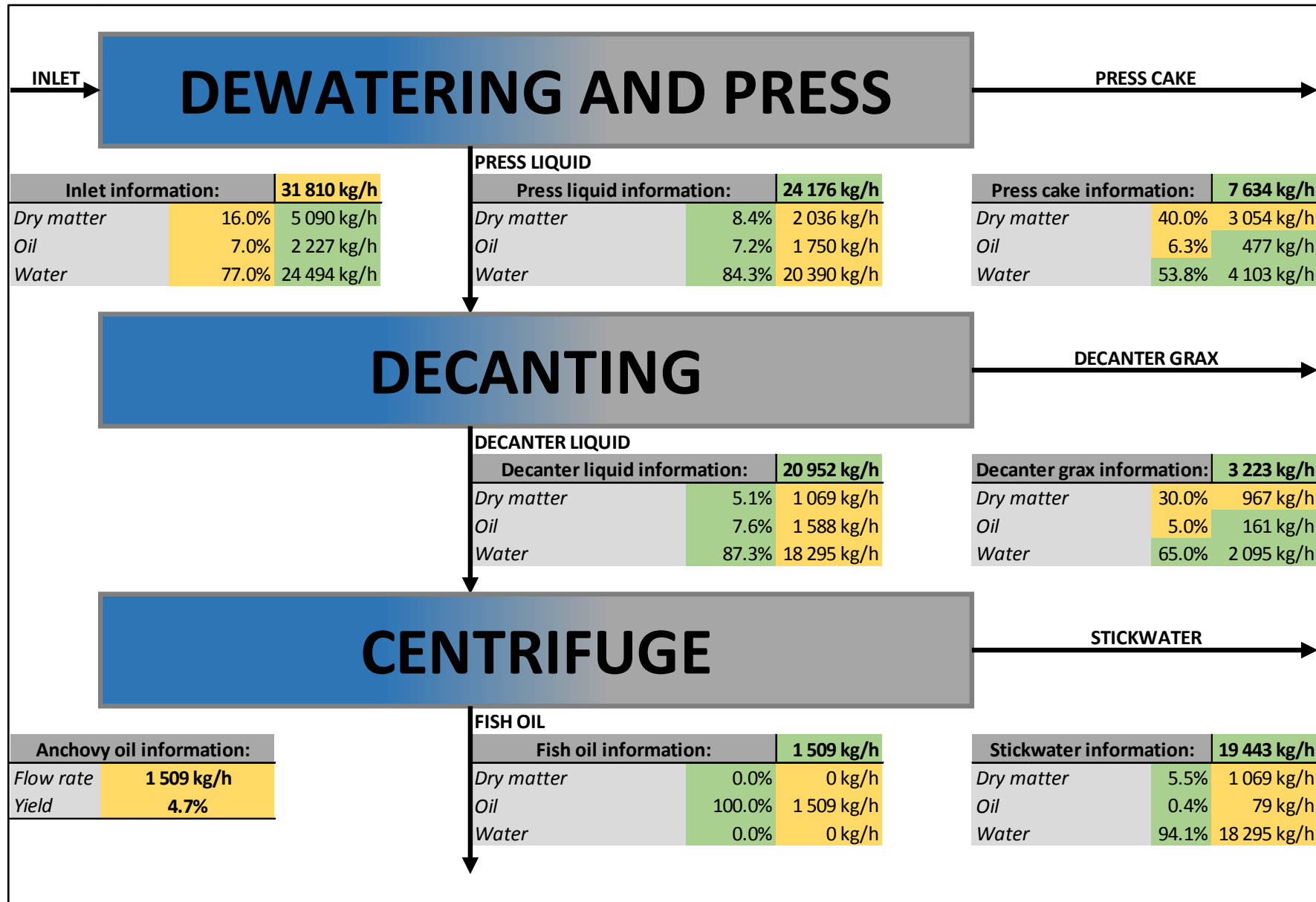
Appendix Figure 5: Simulation of the separation section of the base case process of Factory B in Microsoft Excel 2016.

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Dryer evaporate information:

Flow rate	**8 144 kg/h**

EVAPORATE FROM DRYERS

WASTE HEAT PLANT

STICKWATER

CONCENTRATE

EVAPORATED WATER

Stickwater information:		19 443 kg/h
Dry matter	5.5%	1 069 kg/h
Oil	0.4%	79 kg/h
Water	94.1%	18 295 kg/h

Evaporate information:		15 880 kg/h
Dry matter	0.0%	0 kg/h
Oil	0.0%	0 kg/h
Water	100.0%	15 880 kg/h

Concentrate information:		3 563 kg/h
Dry matter	30.0%	1 069 kg/h
Oil	2.2%	79 kg/h
Water	67.8%	2 414 kg/h
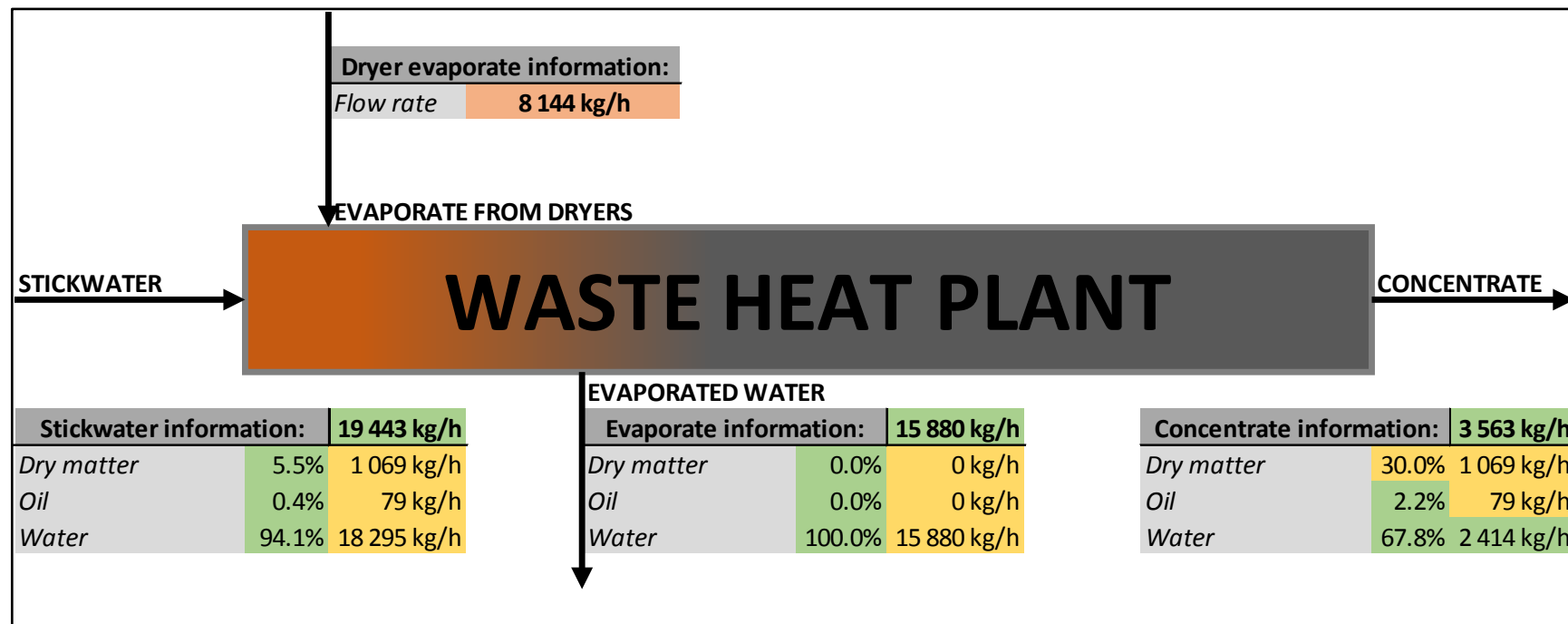
Appendix Figure 6: Simulation of stickwater concentration section ('waste heat plant') of the base case process of Factory B in Microsoft Excel 2016.
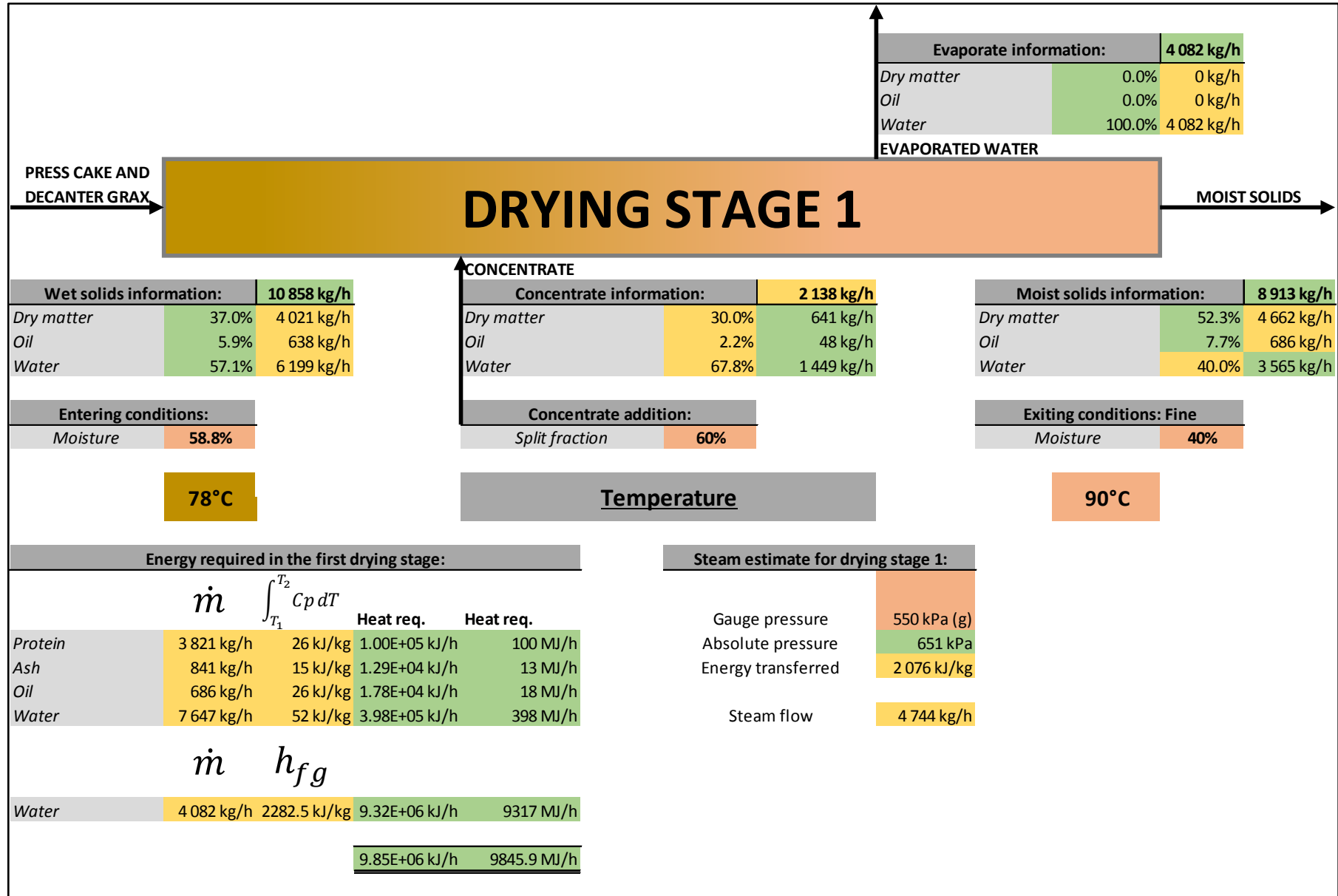
Appendix Figure 7: Simulation of the first drying stage of the base case process of Factory B in Microsoft Excel 2016.

Anchovy-meal information:	
Flow rate	6 276 kg/h
Yield	19.7%
Protein	66.5%
Ash	14.6%
Oil	11.4%
Water	7.5%

Evaporate information:		4 062 kg/h
Dry matter	0.0%	0 kg/h
Oil	0.0%	0 kg/h
Water	100.0%	4 062 kg/h

EVAPORATED WATER

MOIST SOLIDS → **DRYING STAGE 2** → **DRY FISHMEAL**

CONCENTRATE

Moist solids information:		8 913 kg/h
Dry matter	52.3%	4 662 kg/h
Oil	7.7%	686 kg/h
Water	40.0%	3 565 kg/h

Concentrate information:		1 425 kg/h
Dry matter	30.0%	428 kg/h
Oil	2.2%	32 kg/h
Water	67.8%	966 kg/h

Dry fishmeal information:		6 276 kg/h
Dry matter	81.1%	5 090 kg/h
Oil	11.4%	718 kg/h
Water	7.5%	469 kg/h

Entering conditions:	
Moisture	44%

84°C   **Temperature**   **97°C**

Energy required in the second drying stage:				
	$\dot{m}$	$\int_{T_1}^{T_2} Cp\, dT$	Heat req.	Heat req.
Protein	4 171 kg/h	28 kJ/kg	1.18E+05 kJ/h	118 MJ/h
Ash	919 kg/h	17 kJ/kg	1.53E+04 kJ/h	15 MJ/h
Oil	718 kg/h	28 kJ/kg	2.01E+04 kJ/h	20 MJ/h
Water	4 531 kg/h	56 kJ/kg	2.54E+05 kJ/h	254 MJ/h
	$\dot{m}$	h_{fg}		
Water	4 062 kg/h	2264.3 kJ/kg	9.20E+06 kJ/h	9198 MJ/h
			9.61E+06 kJ/h	9605.9 MJ/h

Steam estimate for drying stage 2:	
Gauge pressure	550 kPa (g)
Absolute pressure	651 kPa
Energy transferred	2 076 kJ/kg
Steam flow	4 628 kg/h

Appendix Figure 8: Simulation of the second drying stage of the base case process of Factory B in Microsoft Excel 2016.

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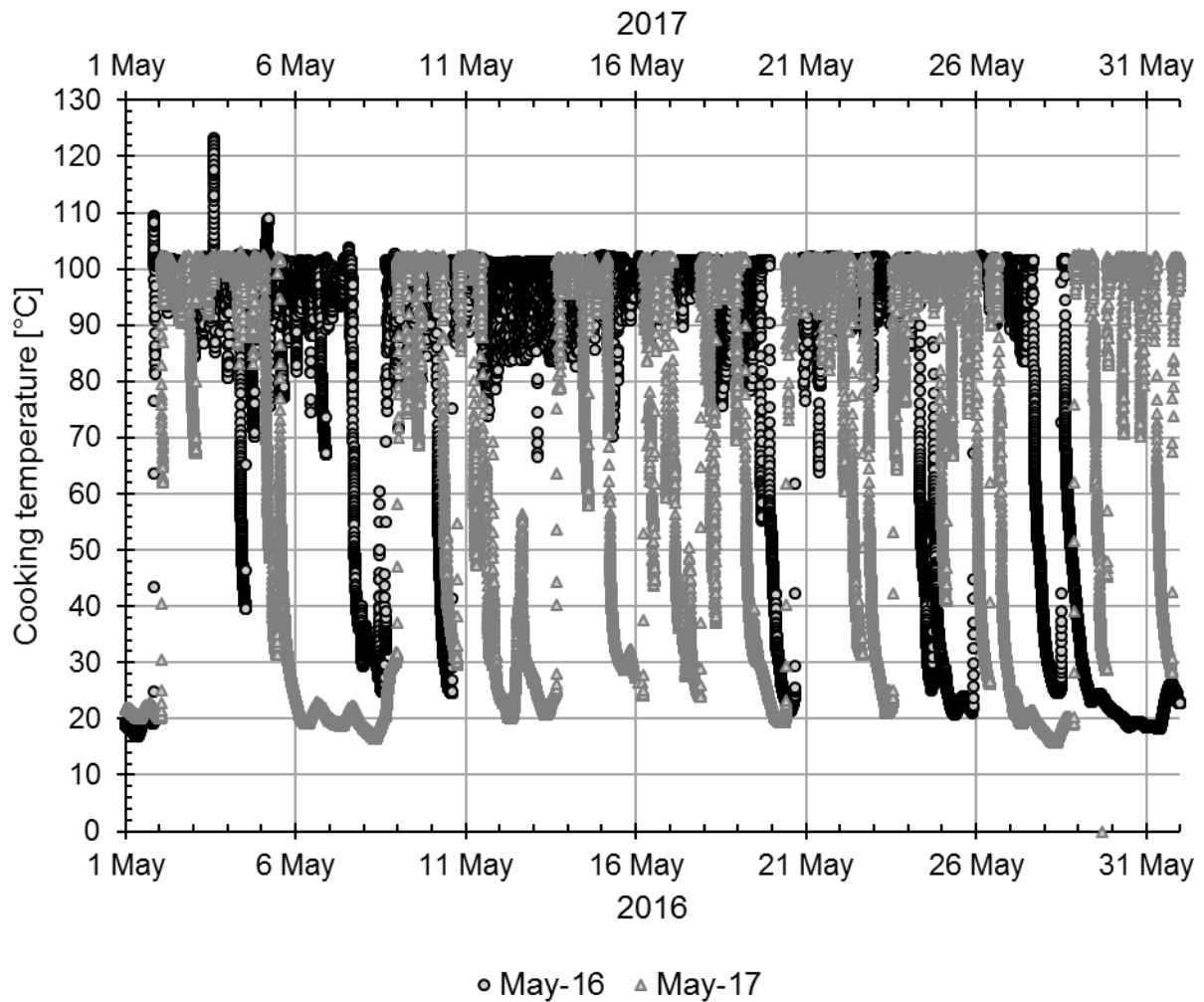
Appendix Figure 9: Recorded outlet temperature of Cooker 3 at Factory B during May of 2016 and 2017. Cooker 3 was selected as reference since it was used for production the most during the period investigated. Since the temperatures were recorded regardless of whether production was taking place, a value of 75°C was defined as the starting temperature of cooking during production. The average of all values >= 75°C was 98.5°C the value used as the outlet temperature of the cooking stage in the base case of Factory B.
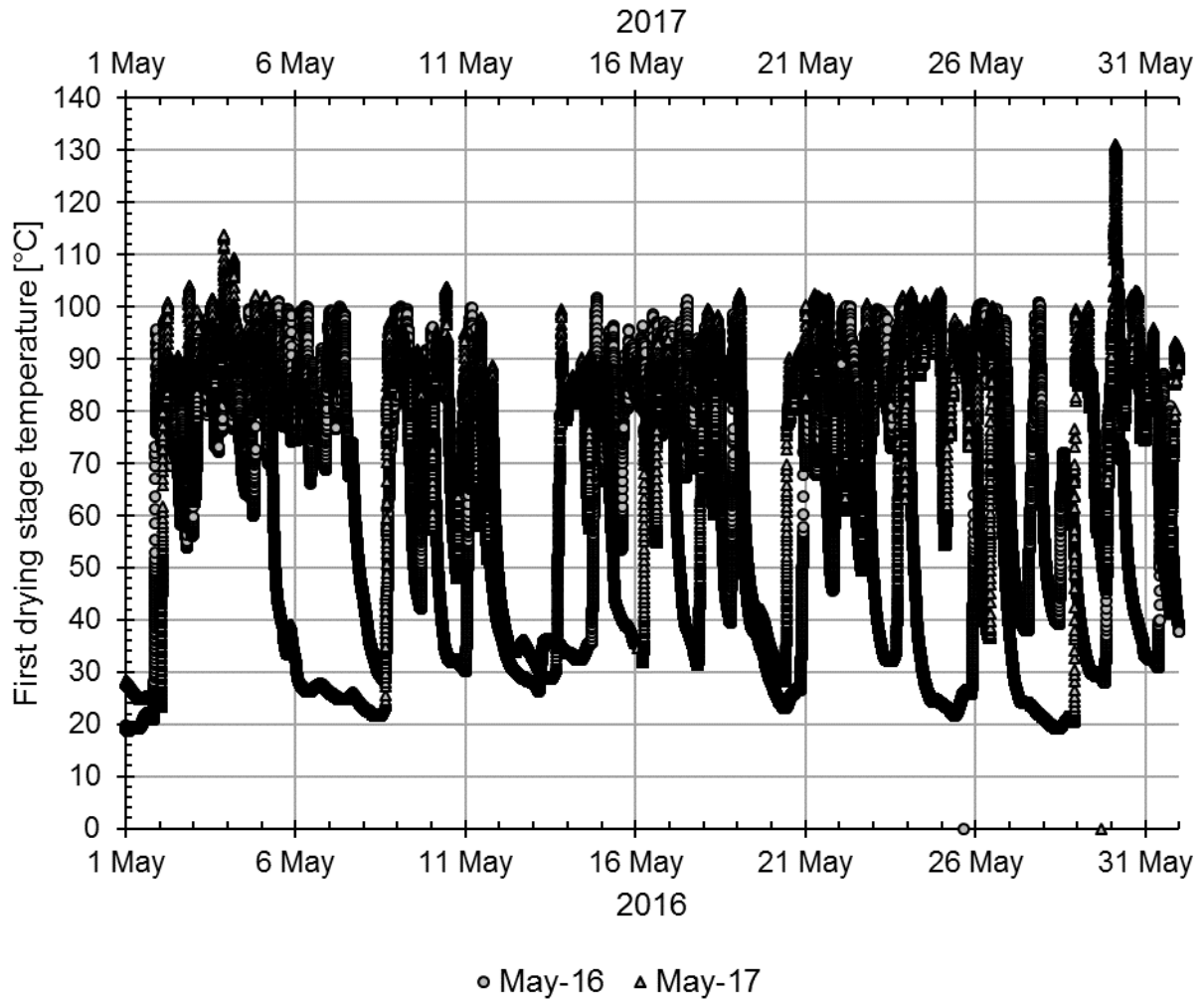
Appendix Figure 10: Recorded outlet temperature of the first drying stage at Factory B during May of 2016 and 2017. Since the temperatures were recorded regardless of whether production was taking place, a value of 80°C was defined as the starting temperature of drying in the first stage during production. The average of all values >= 80°C was 90°C the value used as the outlet temperature of the first drying stage in the base case of Factory B.
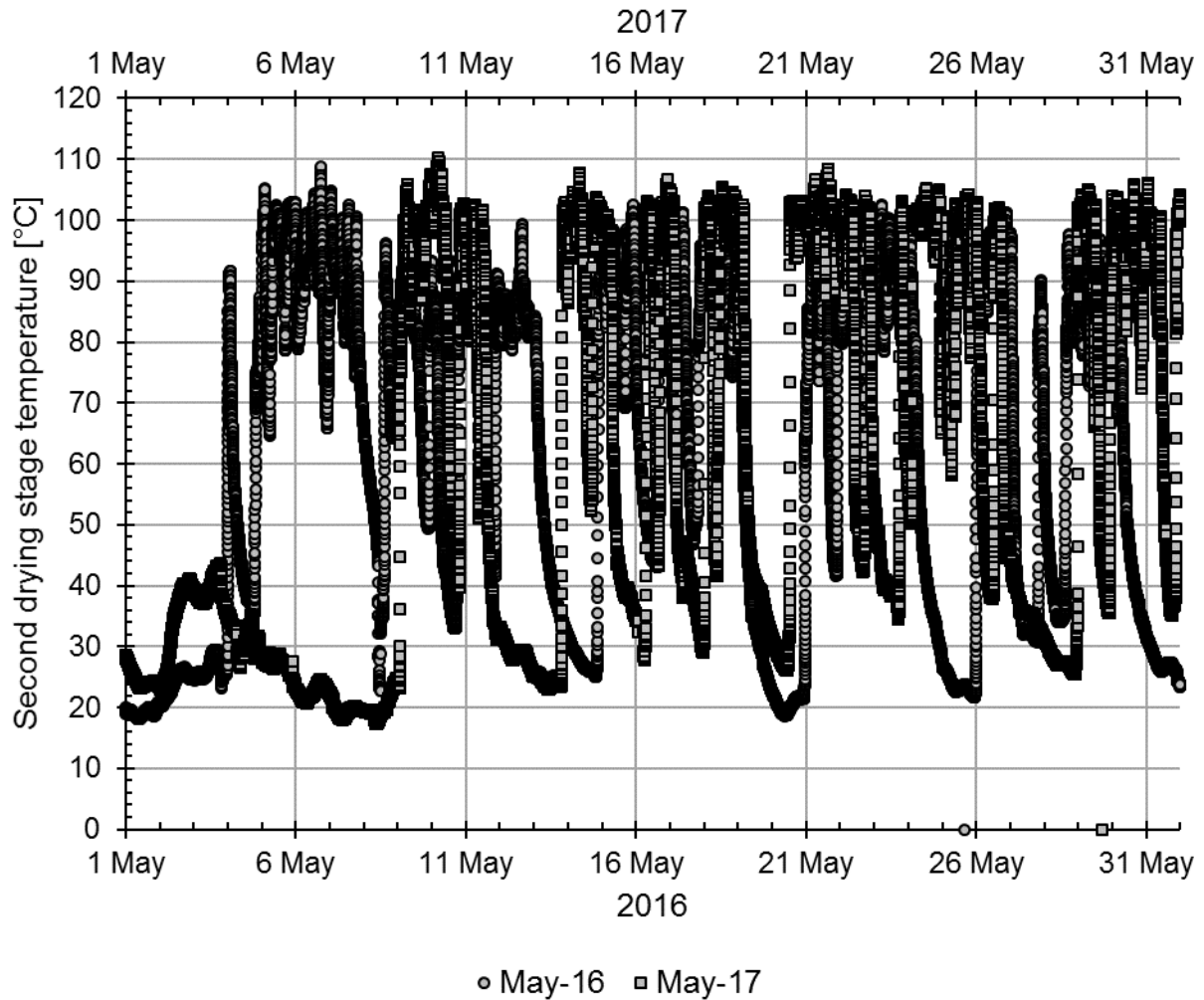
Appendix Figure 11: Recorded outlet temperature of the second drying stage at Factory B during May of 2016 and 2017. Since the temperatures were recorded regardless of whether production was taking place, a value of 80°C was defined as the starting temperature of drying in the second stage during production. The average of all values >= 80°C was 97°C the value used as the outlet temperature of the second drying stage in the base case of Factory B.

Appendix D:Supplement to Factory A solar heat integration and heat recovery study

Appendix Table 6: Typical meteorological year for Factory A's location, generated in Meteonorm 7. GHI and DHI represent the global horizontal irradiation and diffuse horizontal irradiation, respectively. G_NE and D_NE represent the global and diffuse irradiation on a North-East facing plane (46° East of North) with a 30° incline. Uncertainty of yearly values: GHI = 4%, Beam = 7%, G_NE = 4%, T_{air} = 0.5 °C.

Month	GHI [kWh/m²]	DHI [kWh/m²]	Beam [kWh/m²]	G_NE [kWh/m²]	D_NE [kWh/m²]	T_{air} [°C]
January	255	64	272	244	67	21.3
February	205	56	216	206	60	21.4
March	189	45	236	210	51	19.8
April	130	42	162	154	48	17.5
May	93	36	125	120	42	15.1
June	76	29	113	107	35	13.1
July	85	29	129	116	34	12.6
August	110	33	155	139	38	12.9
September	149	50	167	168	55	14.4
October	197	62	205	202	66	16.7
November	232	62	250	224	65	18.6
December	258	74	259	240	75	20.3
TOTAL:	**1982**	**581**	**2286**	**2130**	**636**	**17.0**

Appendix Table 7: Results of the full supply solar heat system for preheating the raw material stream of Factory A. The collector area of the system is 782 m².

Month	Solar heat demand [kWh]	G_NE [kWh/m²]	Solar heat supply [kWh]	|Supply - demand| [kWh]	Auxiliary heat [kWh]	Fuel saved [Litre]
January	36 972	255	85 843	48 871	0	3 610
February	33 611	205	72 474	38 863	0	3 282
March	38 737	189	73 882	35 145	0	3 782
April	36 888	130	54 180	17 292	0	3 602
May	34 703	93	42 218	7 515	0	3 389
June	37 644	76	37 644	0	0	3 676
July	36 888	85	40 811	3 923	0	3 602
August	36 720	110	48 903	12 182	0	3 586
September	37 644	149	59 105	21 461	0	3 676
October	34 872	197	71 067	36 196	0	3 405

Month	Solar heat demand [kWh]	G_NE [kWh/m²]	Solar heat supply [kWh]	|Supply - demand| [kWh]	Auxiliary heat [kWh]	Fuel saved [Litre]
November	36 720	232	78 807	42 087	0	3 586
December	38 905	258	84 436	45 531	0	3 799
TOTAL:	440 306	1 979	749 371	309 065	0	42 994

Appendix Table 8: Results of the option A solar heat system for preheating the raw material stream of Factory A. The collector area of the system is 384 m².

Month	Solar heat demand [kWh]	G_NE [kWh/m²]	Solar heat supply [kWh]	|Supply - demand| [kWh]	Auxiliary heat [kWh]	Fuel saved [Litre]
January	36 972	255	42 122	5 150	0	3 610
February	33 611	205	35 562	1 951	0	3 282
March	38 737	189	36 253	2 484	2 484	3 504
April	36 888	130	26 585	10 303	10 303	2 265
May	34 703	93	20 716	13 988	13 988	1 563
June	37 644	76	18 472	19 173	19 173	1 163
July	36 888	85	20 025	16 863	16 863	1 396
August	36 720	110	23 996	12 724	12 724	1 926
September	37 644	149	29 002	8 642	8 642	2 561
October	34 872	197	34 872	0	0	3 405
November	36 720	232	38 669	1 949	0	3 586
December	38 905	258	41 432	2 527	0	3 799
TOTAL:	440 306	1 979	367 705	95 754	84 178	32 061

Appendix Table 9 Results of the option B solar heat system for preheating the raw material stream of Factory A. The collector area of the system is 337 m².

Month	Solar heat demand [kWh]	G_NE [kWh/m²]	Solar heat supply [kWh]	|Supply - demand| [kWh]	Auxiliary heat [kWh]	Fuel saved [Litre]
January	36 972	255	36 972	0	0	3 610
February	33 611	205	31 214	2 397	2 397	3 004
March	38 737	189	31 820	6 916	6 916	2 903
April	36 888	130	23 335	13 553	13 553	1 833
May	34 703	93	18 183	16 520	16 520	1 226

Month	Solar heat demand [kWh]	G_NE [kWh/m²]	Solar heat supply [kWh]	|Supply - demand| [kWh]	Auxiliary heat [kWh]	Fuel saved [Litre]
June	37 644	76	16 213	21 431	21 431	867
July	36 888	85	17 577	19 311	19 311	1 072
August	36 720	110	21 062	15 658	15 658	1 538
September	37 644	149	25 456	12 188	12 188	2 089
October	34 872	197	30 608	4 263	4 263	2 877
November	36 720	232	33 942	2 778	2 778	3 261
December	38 905	258	36 366	2 539	2 539	3 510
TOTAL:	**440 306**	**1 979**	**322 749**	**117 556**	**117 556**	**27 790**

Appendix Table 10: Base case process results of Factory A when the raw material stream is preheated with the option A solar heat system. The collector area of the system is 384 m².

Month	Production time [hours]	Solar heat supply [MJ]	Available heating rate [MJ/h]	T_{final} [°C]	$\dot{Q}_{boiler}$ [MJ/h]	$\dot{V}_{HFO}$ [litre/h]
January	440	151 639	345	70.0	3 276	92.0
February	400	128 023	320	70.0	3 276	92.0
March	461	130 509	283	66.7	3 297	92.6
April	439	95 707	218	53.2	3 384	95.0
May	413	74 577	181	45.5	3 433	96.4
June	448	66 498	148	38.8	3 475	97.6
July	439	72 091	164	42.1	3 455	97.0
August	437	86 385	198	49.0	3 411	95.8
September	448	104 407	233	56.3	3 364	94.5
October	415	125 537	302	70.0	3 276	92.0
November	437	139 210	319	70.0	3 276	92.0
December	463	149 153	322	70.0	3 276	92.0

Appendix Table 11: Base case process results of Factory A when the raw material stream is preheated with the option B solar heat system. The collector area of the system is 337 m².

Month	Production time [hours]	Solar heat supply [MJ]	Available heating rate [MJ/h]	T_{final} [°C]	$\dot{Q}_{boiler}$ [MJ/h]	$\dot{V}_{HFO}$ [litre/h]
January	440	133 100	303	70.0	3 276	92.0
February	400	112 371	281	66.2	3 300	92.7

Month	Production time [hours]	Solar heat supply [MJ]	Available heating rate [MJ/h]	T_{final} [°C]	$\dot{Q}_{boiler}$ [MJ/h]	$\dot{V}_{HFO}$ [litre/h]
March	461	114 553	248	59.5	3 344	93.9
April	439	84 006	191	47.7	3 419	96.0
May	413	65 459	158	40.9	3 462	97.2
June	448	58 368	130	35.1	3 499	98.2
July	439	63 277	144	37.9	3 481	97.7
August	437	75 823	174	44.0	3 442	96.7
September	448	91 643	205	50.4	3 402	95.5
October	415	110 189	266	63.0	3 321	93.3
November	437	122 190	280	65.9	3 302	92.7
December	463	130 918	283	66.6	3 298	92.6

Appendix E: Supplement to Factory B solar heat integration and heat recovery study

Appendix Table 12: Additional results of the steam production system simulation for the base case process of Factory B and various investigated scenarios.

Scenario	Make-up water required [kg/h]	Condensate recovered [kg/h]	Steam produced [kg/h]	$\dot{Q}_{boiler,required}$ [MJ/h]
Base case	9 459	3 733	13 192	34 728
Fish preheating	7 686	3 047	10 733	28 248
Stickwater concentrate reheating	9 350	3 697	13 047	34 343
Make-up water preheating	9 459	3 733	13 192	32 530
Make-up water and fish preheating	7 686	3 047	10 733	26 463

Appendix Table 13: Effect of raw material preheating on the temperature, energy and steam requirements of the base case process of Factory B.

Section	T_{in} [°C]	T_{out} [°C]	$\dot{Q}_{required}$ [MJ/h]	Source	$\dot{M}_{steam}$ [kg/h]
Preheating	22.0	70.0	5 547.0	Solar	-
Cooking	70.0	98.5	3 317.7	Steam	1 471.2
Drying1	77.5	90.0	9 845.9	Steam	4 688.0
Drying 2	83.5	97.0	9 605.9	Steam	4 573.7

Appendix Table 14: Effect of heating the stickwater concentrate prior to drying, on the temperature, energy and steam requirements of the base case process of Factory B.

Section	T_{in} [°C]	T_{out} [°C]	$\dot{Q}_{required}$ [MJ/h]	Source	$\dot{M}_{steam}$ [kg/h]
Cooking	22.0	98.5	8 864.3	Steam	3 930.7
Drying1	81.8	90.0	9 662.5	Steam	4 600.6
Drying 2	87.6	97.0	9 483.7	Steam	4 515.5

Appendix Table 15: Typical meteorological year for Factory B's location, generated in Meteonorm 7. GHI and DHI represent the global horizontal irradiation and diffuse horizontal irradiation, respectively. G_N and D_N represent the global and diffuse irradiation on a North facing plane with a 35° incline. Uncertainty of yearly values: GHI = 4%, Beam = 8%, G_N = 5%, T_{air} = 0.8 °C.

Month	GHI [kWh/m²]	DHI [kWh/m²]	Beam [kWh/m²]	G_N [kWh/m²]	D_N [kWh/m²]	T_{air} [°C]
January	257	63	275	229	64	22.1
February	207	56	217	206	60	22.1
March	190	48	228	219	57	20.1
April	134	42	166	179	51	17.6

Month	GHI [kWh/m²]	DHI [kWh/m²]	Beam [kWh/m²]	G_N [kWh/m²]	D_N [kWh/m²]	T_{air} [°C]
May	97	36	128	145	44	14.7
June	80	29	119	130	38	11.9
July	91	31	133	144	41	11.3
August	118	38	157	168	48	12
September	153	52	167	186	60	14.6
October	202	60	211	211	65	17.5
November	234	61	255	215	64	19.6
December	261	65	279	225	66	21.1
TOTAL:	**2025**	**582**	**2336**	**2258**	**658**	**17.1**

Appendix Table 16: Solar heat demand [kWh] for the different integration opportunities investigated for Factory B. Fish preheating occurred to 70°C, stickwater concentrate was heated to 75°C and make-up water was preheated to 75°C.

Month	Fish preheating	Stick water concentrate heating	Make-up water preheating	Fish and make-up water preheating
January	0	0	0	0
February	137 124	7 555	54 322	181 251
March	172 561	9 508	68 361	228 091
April	690 244	38 030	273 442	912 365
May	468 380	25 806	185 550	619 105
June	714 895	39 388	283 208	944 949
July	295 819	16 299	117 189	391 014
August	135 584	7 470	53 712	179 215
September	246 516	13 582	97 658	325 845
October	47 762	2 632	18 921	63 132
November	0	0	0	0
December	0	0	0	0

Appendix Table 17: Results for the simulation of fish preheating in Factory B with a 1 751 m² (option A) solar thermal system. January, November and December are not included since no production takes place during these months.

Month	$Q_{available}$ [MJ/h]	T_{out} [°C]	$\dot{Q}_{steam}$ [MJ/h]	$\dot{M}_{steam}$ [kg/h]	$\dot{Q}_{Boiler}$ [MJ/h]	Coal used [kg/h]	$\dot{M}_{fresh}$ [kg/h]
February	6 566	70.0	22 769	10 733	28 248	1 689	7 686
March	5 547	70.0	22 769	10 733	28 248	1 689	7 686
April	1 133	31.8	27 183	12 690	33 405	1 998	9 100
May	1 353	33.7	26 963	12 592	33 148	1 982	9 029

Month	$\dot{Q}_{available}$ [MJ/h]	T_{out} [°C]	$\dot{Q}_{steam}$ [MJ/h]	$\dot{M}_{steam}$ [kg/h]	$\dot{Q}_{Boiler}$ [MJ/h]	Coal used [kg/h]	$\dot{M}_{fresh}$ [kg/h]
June	795	28.9	27 521	12 840	33 800	2 021	9 208
July	2 127	40.5	26 189	12 249	32 244	1 928	8 784
August	5 415	68.9	22 901	10 791	28 405	1 699	7 737
September	3 298	50.6	25 019	11 730	30 881	1 847	8 419
October	19 307	70.0	22 769	10 733	28 248	1 689	7 686

Appendix Table 18: Results for the simulation of fish preheating in Factory B with a 503 m² (option B) solar thermal system. January, November and December are not included since no production takes place during these months.

Month	$\dot{Q}_{available}$ [MJ/h]	T_{out} [°C]	$\dot{Q}_{steam}$ [MJ/h]	$\dot{M}_{steam}$ [kg/h]	$\dot{Q}_{Boiler}$ [MJ/h]	Coal used [kg/h]	$\dot{M}_{fresh}$ [kg/h]
February	1 886	38.4	26 430	12 356	32 525	1 945	8 860
March	1 593	35.8	26 723	12 486	32 870	1 966	8 960
April	326	24.8	27 990	13 048	34 347	2 054	9 355
May	389	25.4	27 927	13 020	34 274	2 050	9 335
June	228	24.0	28 088	13 091	34 461	2 061	9 387
July	611	27.3	27 705	12 921	34 016	2 034	9 269
August	1 556	35.5	26 760	12 502	32 912	1 968	8 966
September	947	30.2	27 369	12 772	33 621	2 011	9 158
October	5 547	70.0	22 769	10 733	28 252	1 689	7 695

Appendix Table 19: Results for the simulation of stickwater concentrate reheating in Factory B with a 96 m² (option A) solar thermal system. January, November and December are not included since no production takes place during these months.

Month	$\dot{Q}_{available}$ [MJ/h]	T_{out} [°C]	$\dot{Q}_{Drying,1}$ [MJ/h]	$\dot{Q}_{Drying,2}$ [MJ/h]	$\dot{Q}_{steam}$ [MJ/h]	$\dot{M}_{steam}$ [kg/h]	$\dot{Q}_{Boiler}$ [MJ/h]	Coal used [kg/h]	$\dot{M}_{fresh}$ [kg/h]
February	362	75.0	9 662	9 484	28 010	13 047	34 343	2 054	9 350
March	306	75.0	9 662	9 484	28 010	13 047	34 343	2 054	9 350
April	62	55.1	9 808	9 581	28 254	13 163	34 649	2 072	9 438
May	75	56.1	9 801	9 576	28 241	13 157	34 634	2 071	9 434
June	44	53.6	9 820	9 588	28 272	13 171	34 673	2 073	9 444
July	117	59.6	9 775	9 559	28 199	13 136	34 581	2 068	9 419
August	298	74.4	9 667	9 487	28 018	13 050	34 353	2 054	9 357
September	182	64.9	9 737	9 533	28 134	13 106	34 500	2 063	9 397
October	1 064	75.0	9 662	9 484	28 010	13 047	34 343	2 054	9 350

Appendix Table 20: Results for the simulation of stickwater concentrate reheating in Factory B with a 28 m² (option B) solar thermal system. January, November and December are not included since no production takes place during these months.

Month	$\dot{Q}_{available}$ [MJ/h]	T_{out} [°C]	$\dot{Q}_{Drying,1}$ [MJ/h]	$\dot{Q}_{Drying,2}$ [MJ/h]	$\dot{Q}_{steam}$ [MJ/h]	$\dot{M}_{steam}$ [kg/h]	$\dot{Q}_{Boiler}$ [MJ/h]	Coal used [kg/h]	$\dot{M}_{fresh}$ [kg/h]
February	104	58.5	9 783	9 564	28 212	13 143	34 597	2 069	9 424
March	88	57.2	9 793	9 571	28 228	13 150	34 617	2 070	9 429
April	18	51.5	9 835	9 599	28 298	13 184	34 705	2 075	9 453
May	21	51.8	9 833	9 597	28 295	13 182	34 701	2 075	9 452
June	13	51.0	9 838	9 601	28 303	13 186	34 712	2 076	9 455
July	34	52.8	9 826	9 592	28 282	13 176	34 685	2 074	9 448
August	86	57.0	9 794	9 572	28 230	13 151	34 620	2 070	9 430
September	52	54.3	9 814	9 585	28 264	13 167	34 662	2 073	9 441

Month	$\dot{Q}_{available}$ [MJ/h]	T_{out} [°C]	$\dot{Q}_{Drying,1}$ [MJ/h]	$\dot{Q}_{Drying,2}$ [MJ/h]	$\dot{Q}_{steam}$ [MJ/h]	$\dot{M}_{steam}$ [kg/h]	$\dot{Q}_{Boiler}$ [MJ/h]	Coal used [kg/h]	$\dot{M}_{fresh}$ [kg/h]
October	306	75.0	9 663	9 484	28 010	13 047	34 344	2 054	9 355

Appendix Table 21: Results for the simulation of make-up water preheating in Factory B with a 694 m² (option A) solar thermal system. January, November and December are not included since no production takes place during these months.

Month	$\dot{Q}_{available}$ [MJ/h]	$T_{Make-up}$ [°C]	$T_{boiler\ inlet}$ [°C]	$\dot{Q}_{Boiler}$ [MJ/h]	Coal used [kg/h]
February	2 601	75.0	86.0	32 530	1 945
March	2 197	75.0	86.0	32 530	1 945
April	449	31.5	55.5	34 279	2 050
May	536	33.7	57.1	34 191	2 045
June	315	28.1	53.1	34 413	2 058
July	843	41.5	62.5	33 885	2 026
August	2 145	73.8	85.1	32 582	1 948
September	1 306	53.1	70.6	33 421	1 999
October	7 649	75.0	86.0	32 530	1 945

Appendix Table 22: Results for the simulation of make-up water preheating in Factory B with a 199 m² (option B) solar thermal system. January, November and December are not included since no production takes place during these months.

Month	$\dot{Q}_{available}$ [MJ/h]	$T_{Make-up}$ [°C]	$T_{boiler\ inlet}$ [°C]	$\dot{Q}_{Boiler}$ [MJ/h]	Coal used [kg/h]
February	747	39.1	60.8	33 980	2 032
March	631	36.2	58.7	34 096	2 039
April	129	23.3	49.8	34 599	2 069
May	154	24.0	50.2	34 574	2 068
June	90	22.3	49.1	34 637	2 071
July	242	26.2	51.8	34 485	2 062
August	616	35.8	58.5	34 111	2 040
September	375	29.6	54.2	34 352	2 054
October	2 197	75.0	86.0	32 530	1 945

Appendix Table 23: Results for the simulation of fish and make-up water preheating in Factory B with a 2 341 m² (option A) solar thermal system. January, November and December are not included since no production takes place during these months.

Month	$\dot{Q}_{available,fish}$ [MJ/h]	$\dot{Q}_{available,make-up}$ [MJ/h]	T_{fish} [°C]	$T_{make-up}$ [°C]	$\dot{Q}_{Boiler}$ [MJ/h]	Coal used [kg/h]
February	6 566	2 113	70.0	75.0	26 463	1 582
March	5 547	1 785	70.0	75.0	26 463	1 582
April	1 133	365	31.8	29.7	33 040	1 976
May	1 353	435	33.7	31.7	32 712	1 956
June	795	256	28.9	26.8	33 544	2 006
July	2 127	685	40.5	38.8	31 560	1 887
August	5 415	1 743	68.9	73.4	26 662	1 594

Month	$\dot{Q}_{available,fish}$ [MJ/h]	$\dot{Q}_{available,make-up}$ [MJ/h]	T_{fish} [°C]	$T_{make-up}$ [°C]	$\dot{Q}_{Boiler}$ [MJ/h]	Coal used [kg/h]
September	3 298	1 061	50.6	50.3	29 819	1 783
October	19 307	6 213	70.0	75.0	26 463	1 582

Appendix Table 24: Results for the simulation of fish and make-up water preheating in Factory B with a 665 m² (option B) solar thermal system. January, November and December are not included since no production takes place during these months.

Month	$\dot{Q}_{available,fish}$ [MJ/h]	$\dot{Q}_{available,make-up}$ [MJ/h]	T_{fish} [°C]	$T_{make-up}$ [°C]	$\dot{Q}_{Boiler}$ [MJ/h]	Coal used [kg/h]
February	1 886	607	38.4	36.6	31 918	1 909
March	1 593	513	35.8	33.9	32 357	1 935
April	326	105	24.8	22.7	34 242	2 048
May	389	125	25.4	23.3	34 149	2 042
June	228	73	24.0	21.9	34 388	2 056
July	611	197	27.3	25.2	33 819	2 022
August	1 556	501	35.5	33.5	32 411	1 938
September	947	305	30.2	28.1	33 317	1 992
October	5 547	1 785	70.0	74.9	26 468	1 583

Appendix F: Supplement to preliminary economic analysis

	Scenario:	Option A (384 m²)			Capital cost	R	3 244 369
	HFO saved (Litre):	32 061			Year 1 fuel savings:	R	237 573
	Cost of HFO:	R 7.41					
	Solar area (m²):	384			IRR:		13.2%
	Specific cost (/m²):	EUR 603.00			LCOH:	R	0.79
	Exchange rate (R/EUR):	R 14.03			NPV:	R	3 275 039
	Inflation:	6%					
	HFO cost increase	10.0%					

Year	Capital cost	Savings income	Annual cash flow	Discounted annual cost	IRR calculation	Cumulated dicsounted cost	Solar heat delivered [kWh]	Discounted solar heat [kWh]
0	R -3 244 369	R -	R -3 244 369	R -3 244 369	R -3 244 369	R -3 244 369	0	0
1	R -	R 237 573	R 237 573	R 224 126	R 209 898	R -3 020 243	356 128	335 970
2	R -	R 261 331	R 261 331	R 232 583	R 203 991	R -2 787 660	356 128	316 953
3	R -	R 287 464	R 287 464	R 241 360	R 198 250	R -2 546 299	356 128	299 012
4	R -	R 316 210	R 316 210	R 250 468	R 192 671	R -2 295 831	356 128	282 087
5	R -	R 347 831	R 347 831	R 259 920	R 187 248	R -2 035 912	356 128	266 120
6	R -	R 382 614	R 382 614	R 269 728	R 181 979	R -1 766 184	356 128	251 056
7	R -	R 420 876	R 420 876	R 279 906	R 176 858	R -1 486 277	356 128	236 845
8	R -	R 462 963	R 462 963	R 290 469	R 171 880	R -1 195 808	356 128	223 439
9	R -	R 509 260	R 509 260	R 301 430	R 167 043	R -894 378	356 128	210 792
10	R -	R 560 186	R 560 186	R 312 805	R 162 342	R -581 574	356 128	198 860
11	R -	R 616 204	R 616 204	R 324 609	R 157 774	R -256 965	356 128	187 604
12	R -	R 677 825	R 677 825	R 336 858	R 153 333	R 79 893	356 128	176 985
13	R -	R 745 607	R 745 607	R 349 570	R 149 018	R 429 463	356 128	166 967
14	R -	R 820 168	R 820 168	R 362 761	R 144 824	R 792 224	356 128	157 516
15	R -	R 902 185	R 902 185	R 376 450	R 140 749	R 1 168 674	356 128	148 600
16	R -	R 992 403	R 992 403	R 390 656	R 136 788	R 1 559 330	356 128	140 188
17	R -	R 1 091 643	R 1 091 643	R 405 397	R 132 938	R 1 964 727	356 128	132 253
18	R -	R 1 200 808	R 1 200 808	R 420 695	R 129 197	R 2 385 423	356 128	124 767
19	R -	R 1 320 888	R 1 320 888	R 436 571	R 125 561	R 2 821 993	356 128	117 705
20	R -	R 1 452 977	R 1 452 977	R 453 045	R 122 028	R 3 275 039	356 128	111 042

Appendix Figure 12: Net present value calculation for preheating the raw material stream in Factory A with a 384 m² solar heating system. The SF of this system is 0.81.

Scenario:	Heat recovery
HFO saved (Litre):	6 694
Cost of HFO:	R 7.41
Specific cost (/m^2):	EUR 603.00
Exchange rate (R/EUR):	R 14.03
Inflation:	6%
HFO cost increase	10.0%

Total capital cost:	R	1 122 008
Year 1 fuel savings:	R	49 602
IRR:		7.8%
LCOH:	R	0.22
NPV:	R	239 164

Year	Capital cost	Savings income	Annual cash flow	Discounted annual cost	IRR calculation	Cumulated dicsounted cost	Heat delivered [kWh]	Discounted heat [kWh]
0	R -1 122 008	R -	R -1 122 008	R -1 122 008	R -1 122 008	R -1 122 008	0	
1	R -	R 49 602	R 49 602	R 46 795	R 46 011	R -1 075 213	440 306	415 383
2	R -	R 54 563	R 54 563	R 48 561	R 46 948	R -1 026 653	440 306	391 870
3	R -	R 60 019	R 60 019	R 50 393	R 47 904	R -976 260	440 306	369 689
4	R -	R 66 021	R 66 021	R 52 295	R 48 880	R -923 965	440 306	348 763
5	R -	R 72 623	R 72 623	R 54 268	R 49 875	R -869 697	440 306	329 022
6	R -	R 79 885	R 79 885	R 56 316	R 50 891	R -813 381	440 306	310 398
7	R -	R 87 874	R 87 874	R 58 441	R 51 927	R -754 940	440 306	292 828
8	R -	R 96 661	R 96 661	R 60 646	R 52 984	R -694 294	440 306	276 253
9	R -	R 106 327	R 106 327	R 62 935	R 54 063	R -631 359	440 306	260 616
10	R -	R 116 960	R 116 960	R 65 310	R 55 164	R -566 049	440 306	245 864
11	R -	R 128 656	R 128 656	R 67 774	R 56 288	R -498 275	440 306	231 947
12	R -	R 141 521	R 141 521	R 70 332	R 57 434	R -427 943	440 306	218 818
13	R -	R 155 674	R 155 674	R 72 986	R 58 603	R -354 957	440 306	206 432
14	R -	R 171 241	R 171 241	R 75 740	R 59 797	R -279 217	440 306	194 748
15	R -	R 188 365	R 188 365	R 78 598	R 61 014	R -200 619	440 306	183 724
16	R -	R 207 202	R 207 202	R 81 564	R 62 257	R -119 055	440 306	173 325
17	R -	R 227 922	R 227 922	R 84 642	R 63 525	R -34 413	440 306	163 514
18	R -	R 250 714	R 250 714	R 87 836	R 64 818	R 53 423	440 306	154 258
19	R -	R 275 785	R 275 785	R 91 151	R 66 138	R 144 574	440 306	145 527
20	R -	R 303 364	R 303 364	R 94 590	R 67 485	R 239 164	440 306	137 289

Appendix Figure 13: Net present value calculation for preheating the raw material stream in Factory A with heat recovery from the condensate stream exiting the dryer.

Scenario:	Raw material preheating
Coal saved (MeT):	92
Cost of coal:	R 998.81
Solar area (m^2):	503
Specific cost (/m^2):	EUR 603.00
Exchange rate (R/EUR):	R 15.20
Inflation:	6%
Coal price increase:	8.80%

Capital cost:	R	4 610 542
Year 1 fuel savings:	R	91 745
IRR:		0.0%
NPV:	R	-2 367 843

Year	Capital cost	Savings income	Annual cash flow	Discounted annual cost	IRR calculation
0	R -4 610 542	R -	R -4 610 542	R -4 610 542	R -4 610 542
1	R -	R 91 745	R 91 745	R 86 552	R 91 776
2	R -	R 99 818	R 99 818	R 88 838	R 99 887
3	R -	R 108 602	R 108 602	R 91 185	R 108 715
4	R -	R 118 159	R 118 159	R 93 593	R 118 322
5	R -	R 128 557	R 128 557	R 96 065	R 128 779
6	R -	R 139 870	R 139 870	R 98 603	R 140 160
7	R -	R 152 179	R 152 179	R 101 208	R 152 547
8	R -	R 165 571	R 165 571	R 103 881	R 166 028
9	R -	R 180 141	R 180 141	R 106 625	R 180 701
10	R -	R 195 993	R 195 993	R 109 442	R 196 670
11	R -	R 213 241	R 213 241	R 112 333	R 214 051
12	R -	R 232 006	R 232 006	R 115 300	R 232 968
13	R -	R 252 422	R 252 422	R 118 345	R 253 557
14	R -	R 274 636	R 274 636	R 121 472	R 275 965
15	R -	R 298 804	R 298 804	R 124 680	R 300 353
16	R -	R 325 098	R 325 098	R 127 974	R 326 897
17	R -	R 353 707	R 353 707	R 131 354	R 355 787
18	R -	R 384 833	R 384 833	R 134 824	R 387 230
19	R -	R 418 698	R 418 698	R 138 385	R 421 451
20	R -	R 455 544	R 455 544	R 142 041	R 458 697

Appendix Figure 14: Net present value calculation for preheating the raw material stream in Factory B with a 503 m^2 solar heating system. The SF of this system is 0.12.

Scenario:	Concentrate preheating
Coal saved (MeT):	7
Cost of coal:	R 998.81
Solar area (m²):	28
Specific cost (/m²):	EUR 603.00
Exchange rate (R/EUR):	R 15.20
Inflation:	6%
Coal price increase:	8.80%

Capital cost:	R	254 026
Year 1 fuel savings:	R	6 879
IRR:		2.4%
NPV:	R	-85 874

Year	Capital cost		Savings income		Annual cash flow		Discounted annual cost		IRR calculation	
0	R	-254 026	R	-	R	-254 026	R	-254 026	R	-254 026
1	R	-	R	6 879	R	6 879	R	6 489	R	6 718
2	R	-	R	7 484	R	7 484	R	6 661	R	7 138
3	R	-	R	8 143	R	8 143	R	6 837	R	7 585
4	R	-	R	8 859	R	8 859	R	7 017	R	8 059
5	R	-	R	9 639	R	9 639	R	7 203	R	8 563
6	R	-	R	10 487	R	10 487	R	7 393	R	9 099
7	R	-	R	11 410	R	11 410	R	7 588	R	9 668
8	R	-	R	12 414	R	12 414	R	7 789	R	10 273
9	R	-	R	13 507	R	13 507	R	7 995	R	10 916
10	R	-	R	14 695	R	14 695	R	8 206	R	11 599
11	R	-	R	15 988	R	15 988	R	8 422	R	12 324
12	R	-	R	17 395	R	17 395	R	8 645	R	13 095
13	R	-	R	18 926	R	18 926	R	8 873	R	13 914
14	R	-	R	20 592	R	20 592	R	9 108	R	14 785
15	R	-	R	22 404	R	22 404	R	9 348	R	15 710
16	R	-	R	24 375	R	24 375	R	9 595	R	16 693
17	R	-	R	26 520	R	26 520	R	9 849	R	17 737
18	R	-	R	28 854	R	28 854	R	10 109	R	18 846
19	R	-	R	31 393	R	31 393	R	10 376	R	20 025
20	R	-	R	34 156	R	34 156	R	10 650	R	21 278

Appendix Figure 15: Net present value calculation for preheating the stickwater concentrate in Factory B with a 28 m² solar heating system. The SF of this system is 0.12.

Scenario:	Boiler water preheating
Coal saved (MeT):	32
Cost of coal:	R 998.81
Solar area (m^2):	199
Specific cost (/m^2):	EUR 603.00
Exchange rate (R/EUR):	R 15.20
Inflation:	6%
Coal price increase:	8.80%

Capital cost:	R	1 826 480
Year 1 fuel savings:	R	32 162
IRR:		-0.9%
NPV:	R	-1 040 288

Year	Capital cost		Savings income		Annual cash flow		Discounted annual cost		IRR calculation	
0	R	-1 826 480	R	-	R	-1 826 480	R	-1 826 480	R	-1 826 480
1	R	-	R	32 162	R	32 162	R	30 341	R	32 469
2	R	-	R	34 992	R	34 992	R	31 143	R	35 665
3	R	-	R	38 071	R	38 071	R	31 965	R	39 175
4	R	-	R	41 421	R	41 421	R	32 810	R	43 030
5	R	-	R	45 067	R	45 067	R	33 676	R	47 265
6	R	-	R	49 032	R	49 032	R	34 566	R	51 916
7	R	-	R	53 347	R	53 347	R	35 479	R	57 025
8	R	-	R	58 042	R	58 042	R	36 416	R	62 637
9	R	-	R	63 150	R	63 150	R	37 378	R	68 801
10	R	-	R	68 707	R	68 707	R	38 365	R	75 572
11	R	-	R	74 753	R	74 753	R	39 379	R	83 009
12	R	-	R	81 331	R	81 331	R	40 419	R	91 179
13	R	-	R	88 488	R	88 488	R	41 487	R	100 152
14	R	-	R	96 275	R	96 275	R	42 583	R	110 008
15	R	-	R	104 747	R	104 747	R	43 707	R	120 834
16	R	-	R	113 965	R	113 965	R	44 862	R	132 725
17	R	-	R	123 994	R	123 994	R	46 047	R	145 787
18	R	-	R	134 906	R	134 906	R	47 263	R	160 134
19	R	-	R	146 777	R	146 777	R	48 512	R	175 893
20	R	-	R	159 694	R	159 694	R	49 793	R	193 203

Appendix Figure 16: Net present value calculation for preheating the boiler make-up water stream in Factory B with a 199 m^2 solar heating system. The SF of this system is 0.12.

Scenario:	Combined system
Coal saved (MeT):	117
Cost of coal:	R 998.81
Solar area (m^2):	665
Specific cost (/m^2):	EUR 603.00
Exchange rate (R/EUR):	R 15.20
Inflation:	6%
Coal price increase:	8.80%

Capital cost:	R	6 094 195
Year 1 fuel savings:	R	116 617
IRR:		-0.3%
NPV:	R	-3 243 499

Year	Capital cost	Savings income	Annual cash flow	Discounted annual cost	IRR calculation
0	R -6 094 195	R -	R -6 094 195	R -6 094 195	R -6 094 195
1	R -	R 116 617	R 116 617	R 110 016	R 117 002
2	R -	R 126 879	R 126 879	R 112 922	R 127 719
3	R -	R 138 044	R 138 044	R 115 905	R 139 418
4	R -	R 150 192	R 150 192	R 118 966	R 152 188
5	R -	R 163 409	R 163 409	R 122 109	R 166 128
6	R -	R 177 789	R 177 789	R 125 334	R 181 345
7	R -	R 193 435	R 193 435	R 128 645	R 197 956
8	R -	R 210 457	R 210 457	R 132 043	R 216 088
9	R -	R 228 977	R 228 977	R 135 531	R 235 881
10	R -	R 249 127	R 249 127	R 139 111	R 257 487
11	R -	R 271 050	R 271 050	R 142 786	R 281 072
12	R -	R 294 903	R 294 903	R 146 558	R 306 817
13	R -	R 320 854	R 320 854	R 150 429	R 334 920
14	R -	R 349 090	R 349 090	R 154 403	R 365 598
15	R -	R 379 809	R 379 809	R 158 481	R 399 086
16	R -	R 413 233	R 413 233	R 162 667	R 435 641
17	R -	R 449 597	R 449 597	R 166 964	R 475 544
18	R -	R 489 162	R 489 162	R 171 375	R 519 102
19	R -	R 532 208	R 532 208	R 175 902	R 566 650
20	R -	R 579 042	R 579 042	R 180 548	R 618 554

Appendix Figure 17: Net present value calculation for preheating the boiler make-up water and entering fish stream in Factory B with a 665 m^2 solar heating system. The SF of this system is 0.12.

Appendix Table 25: Results of the investigation into the effect of fuel type and solar fraction into the net present value of the fish preheating solar heat systems proposed for Factory B.

Solar fraction	Total collector area [m²]	Total coal saved [tons]	Coal based NPV	Possible HFO Saved [litre]	HFO based NPV	Total solar heat [kWh]	Excess solar heat [kWh]
0.05	204	38	R -935 922	26 459	R 3 514 778	206 717	61 273
0.12	503	92	R -2 367 843	63 836	R 8 370 019	510 899	151 436
0.20	862	148	R -4 292 457	102 654	R 12 974 967	875 287	293 510
0.30	1 331	221	R -6 808 978	153 495	R 19 010 514	1 352 074	479 409
0.39	1 751	280	R -9 603 547	194 532	R 23 507 583	1 778 412	670 832
0.50	2 725	367	R -16 014 846	255 099	R 26 895 504	2 767 735	1 313 293
0.75	5 611	550	R -38 001 633	382 134	R 26 277 286	5 698 666	3 517 003
0.90	8 013	660	R -57 343 299	458 370	R 19 759 278	8 138 726	5 520 729
1.00	12 220	733	R -94 108 486	509 477	R -8 409 155	12 411 678	9 502 795